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Effect of SO2 Emission Requirements on Fluidized-bed Combustion Systems: Preliminary Technical/Economic Assessment

Westinghouse Research and Development Center, Pittsburgh, PA

Prepared for

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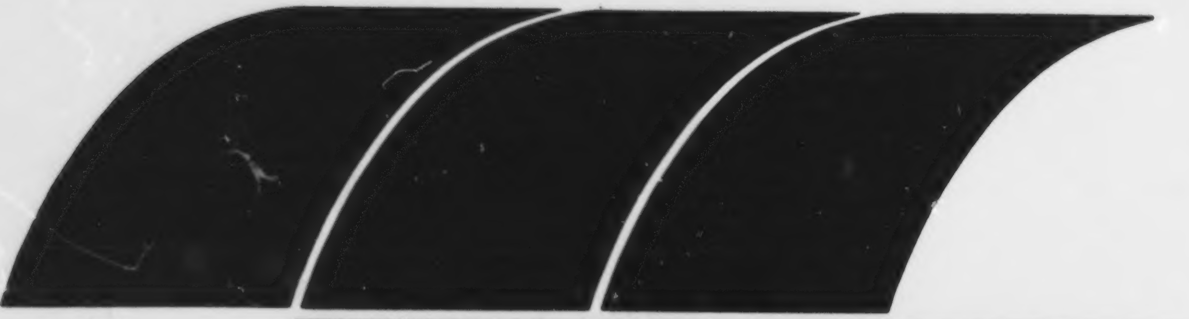

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16. ABSTRACT

The report gives results of a preliminary technical/economic evaluation to project the impact of SO2 control requirements (up to 90% control) on the capital and energy costs of atmospheric-pressure and pressurized fluidized-bed combustion (AFBC and PFBC) power plants. Ability of AFBC and PFBC to reduce emissions of particulates and NOx is also considered. Performance and economic projections are presented, both for the current New Source Performance Standards and for more stringent controls; the AFBC and PFBC projections are compared with equivalent projections for conventional boilers with flue gas desulfurization. The projections show that AFBC and PFBC plants can achieve SO2 control up to at least 90%, and still remain economically competitive with conventional boilers using scrubbers. However, FBC plant design and operating parameters are critical to the achievement of high levels of SO2 control at competitive energy costs. In particular, increased gas residence time in the bed, and reduced sorbent particle size, are important in effective attainment of high levels of SO2 control. The projections of FBC performance at 90% SO2 removal must be confirmed experimentally on FBC units large enough to be representative of commercial-scale combustors.

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Fluidized Bed Processing	Boilers	Stationary Sources	13H, 07A	13A
Sulfur Dioxide	Flue Gases	Energy Requirements	07B	21B
Capitalized Costs	Desulfurization	Particulates	14A, 05A	07D
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PREFACE

The Westinghouse R&D Center is carrying out a program to provide experimental and engineering support for the development of fluidized-bed combustion systems under contract to the Industrial Environmental Research Laboratory, U. S. Environmental Protection Agency (EPA), at Research Triangle Park, NC. The contract scope includes atmospheric and pressurized fluidized-bed combustion processes as they may be applied for steam generation, electric power generation, or process heat. Specific tasks include work on calcium-based sulfur removal systems (e.g., sorption kinetics, regeneration, attrition, modeling), alternative sulfur sorbents, nitrogen oxide emissions, particulate emissions and control, trace element emissions and control, spent sorbent and ash disposal, and systems evaluation (e.g., impact of new source performance standards on fluidized-bed combustion system design and cost).

This document contains the results of a brief technical effort, defined and completed under a technical directive, from August 1977 to January 1978. The work reported represents a preliminary application of prior work completed by Westinghouse under contract to EPA. This prior work on fluidized-bed combustion is found in references 1 through 7.

This study was conducted in close coordination with EPA's Office of Energy, Minerals and Industry headquarters office; EPA's Office of Planning and Evaluation; and a related study for OPE by Energy Resources Co. of the appropriateness and differential effects of alternative regulatory approaches to the standards-setting process. Westinghouse provided them with cost and performance data and a methodology for performing FBC parametric studies.

ABSTRACT

An evaluation has been performed to project the impact of up to 90 percent SO_2 control on the capital and energy costs of atmospheric-pressure fluidized-bed combustion (AFBC) and pressurized fluidized-bed combustion (PFBC) power plants. The ability of AFBC and PFBC to achieve reduced emissions of particulates and nitrogen oxides are also considered. Performance and economic projections are presented for current emission standards and for a set of hypothetical emission requirements representing a more stringent degree of control, and are compared with equivalent projections for conventional boilers with flue gas desulfurization. The projections of fluidized-bed combustion performance and system cost show that AFBC and PFBC plants can achieve up to 90 percent SO_2 control and still remain economically competitive with conventional plants with flue gas scrubbers. The selection of fluidized-bed plant design and operating parameters, however, is critical to the achievement of high levels of control at competitive energy costs. The projections of FBC performance at up to 90 percent SO_2 control must be confirmed in the future on FBC units sufficiently large to be representative of commercial-scale combustors.

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ABBREVIATIONS AND SYMBOLS

EPA	Environmental Protection Agency
OPE	Office of Planning and Evaluation
FBC	Fluid-bed combustion
AFBC	Atmospheric fluid-bed combustion
PFBC	Pressurized fluid-bed combustion
NSPS	New source performance standards
TGA	Thermogravimetric analysis
FGD	Flue gas desulfurization
TVA	Tennessee Valley Authority
ECAS	Energy Conservation Alternatives Study
TG	Thermogravimetric
B&W	Babcock and Wilcox
ANL	Argonne National Laboratories
NCB	National Coal Board
FB	Fluid bed
PER	Pope, Evans and Robbins
NASA	National Aeronautics and Space Administration
NSF	National Science Foundation
DOE	Department of Energy
GE	General Electric
EST	Electrostatic Precipitator
BOP	Balance of Plant
A/E	Architect/Engineer

APPENDIX A

h	bed height, expanded
R	Sulfur removal, fractional
$\overset{Z}{\xi}$	gas residence time, expanded bed height/interstitial gas velocity
K	average rate constant for SO_2 sorption in the bed
K^1	first order rate constant for the reaction $CaO + SO_2 + 1/2\ O_2 \longrightarrow CaSO_4$
γ	volume fraction of bed bubbles
ε	bed voidage in emulsion phase
F	fraction of emulsion volume occupied by inerts
U	sorbent utilization, fractional
C	mole SO_2/cc in TG reaction gas
P	solids density, mole Ca/cc
ε_{TG}	TG voidage
C	Ca/S molar feed ratio
d	particle diameter

APPENDIX C

E	coal energy input rate
F_c	coal rate
H	coal heating value, as received
F_s	fresh sorbent feed rate
X_s	coal sulfur content
M	sorbent weight per mole of calcium
F_d	rate of spent solids
X_{CO_2}	weight fraction of sorbent released as CO_2 in calcination
X_A	coal ash content, weight fraction
P	plant power output
P_b	base net power
η	gross generation efficiency
f	combustion efficiency for parametric plant
f_b	combustion efficiency for base plant
H_R	sorbent heat of calcination
$F_{s,b}$	fresh sorbent feed for base plant
ΔP_A	change in auxiliary power between base plant and parametric plant
E_c	energy cost, mills, kWh
p	coal cost (\$/MBtu)
S	sorbent cost (\$/MBtu)
I	capital investment

I. INTRODUCTION

Current fluidized-bed combustion (FBC) power plant development programs, and FBC power plant engineering studies, are focusing attention on the achievement of current new source performance standards (NSPS) for sulfur oxides, particulates, and oxides of nitrogen. In addition, the FBC design philosophy expressed in these programs is generally directed toward minimizing the fluid-bed combustor cost by applying compact designs resulting from the use of high fluidizing velocities and correspondingly large sorbent particle diameters.

Under the 1977 Amendments to the Clean Air Act, periodic revisions to the emission standards will be evaluated by EPA. It is appropriate, therefore, that the impact of potentially more stringent standards on FBC power plant design and economics be addressed in development programs. The relative merits of minimizing the fluid-bed combustor cost versus maximizing its performance (e.g., maximum combustion efficiency, minimum calcium-to-sulfur ratio, minimum particle attrition) must be assessed. Although compact design could reduce combustor cost to a minimum, the resulting plant energy costs could be higher than optimal and the potential for economically meeting more stringent environmental emission goals could be lower.

The objective of this evaluation is to project the impact of more stringent emission requirements on FBC power plant economics, with emphasis on achieving up to 90 percent SO_2 control. The impact of fluid-bed combustor design and operating conditions on FBC environmental control capabilities and on plant economics is assessed. The projected cost of achieving both current standards and more stringent degrees of control is compared with the estimated cost of achieving equivalent degrees of control in a conventional coal boiler equipped with a flue gas scrubber.

The current EPA new source performance standards for large coal-fired boilers are: sulfur oxides, 1.2 lb SO_2 MBtu (516 ng/J); particulates, 0.1 lb/MBtu (43.0 ng/J); and oxides of nitrogen, 0.7 lb NO_2 MBtu (301 ng/J). The more stringent degrees of control considered in this study are:

- Sulfur oxides - 90 percent removal of coal sulfur content
- Particulates - 0.03 lb/MBtu (12.9 ng/J)
- Oxides of nitrogen - 0.6 lb/MBtu (258 ng/J).

The selection of these control/emission levels as the "more stringent degree of control" should not be construed as suggesting that these are the most stringent levels that FBC can achieve or that FBC would ever be required to achieve. These levels have been selected for the current study only because they are one set of values that were considered during the planned revision of the NSPS for utility boilers.

The evaluation incorporates and utilizes the results of previous FBC engineering studies and modeling studies in developing preliminary performance and cost projections for atmospheric-pressure fluidized-bed combustion (AFBC) and pressurized fluidized-bed combustion (PFBC). This work has been performed on a technical directive from the Industrial Environmental Research Laboratory, U. S. Environmental Protection Agency (EPA), during the period of August 1977 to January 1978.

II. SUMMARY

Projections of AFBC and PFBC power plant performance and economics have been developed through the assimilation of previous FBC power plant design studies, FBC performance models, and data assessments. The key parameters in the evaluation are the sorbent calcium-to-sulfur ratio, the coal sulfur content and the fluid-bed combustor design and operating conditions.

The projections of FBC power plant energy costs indicate that--for both the existing SO_2 emission standard and for 90 percent sulfur removal--FBC has considerable potential for being cost competitive with conventional coal-fired power plants using lime slurry scrubbing. The competitiveness of FBC depends upon proper selection of fluid-bed combustor operating conditions--i.e., sufficiently long gas residence time in the bed (sufficiently low gas velocity, and sufficiently deep beds) and sufficiently small sorbent particle size. This selection of variables will result in a less compact combustor, but the cost savings resulting from decreased sorbent requirements would more than compensate for increased combustor costs.

In the design of FBC power plants emphasis should be placed on maximizing the fluid-bed combustor performance rather than on minimizing the combustor cost through compact design. The combustor cost represents a small portion of the FBC power plant investment and is also relatively insensitive to changes in design and operating conditions. On the other hand, the overall FBC power plant cost of electricity is strongly dependent on the combustor performance.

The calcium-to-sulfur molar ratio--that is, the moles of sorbent calcium fed to the fluid-bed combustor divided by the moles of sulfur fed in the coal--is the single, most important performance factor relative to FBC power plant cost and performance for high-sulfur Eastern coals (2 to 5 wt% sulfur). The calcium-to-sulfur ratio has a dramatic impact on the FBC power plant thermal efficiency, capital investment, and cost of electricity. Increased calcium-to-sulfur ratio, if required for lower sulfur oxide emissions, results in increased auxiliary power consumption for solids handling and significant sorbent calcination energy losses. The resulting reduced net plant efficiency and slightly increased equipment costs for solids handling, crushing, drying, feeding, and spent solids disposal lead to increased capital investment and energy costs. In addition, the increased cost of raw sorbent at increased feed rates significantly increases the energy cost.

The all-important projection of sorbent feed requirements was accomplished in this study using a kinetic model for SO_2 capture developed by Westinghouse.[3] This model--using rate constants measured in laboratory thermogravimetric analysis (TGA) equipment, and confirmed where possible using available data from experimental fluidized-bed combustors--is capable of projecting sorbent requirements, where TGA data have been generated, as a function of key combustor operating/design conditions. Further confirmation of the accuracy of model projections is necessary, especially at high levels of SO_2 removal, where very little combustor data are currently available; data from larger combustors, more representative of commercial-scale units, are also necessary to confirm model projections.

While cost and performance uncertainties exist for several subsystems in the FBC power plants (for example, solids feeding and particulate control), these are expected to be resolved through proper design and specification of materials and operating conditions and maintenance and operating procedures. The overall financial impact of these cost/performance uncertainties will probably be small relative to the uncertainties in such site factors as sorbent availability, sorbent cost, coal cost, solid waste disposal feasibility or utilization markets, local emission standards, and so on.

For low-sulfur Western coals and lignites the impact of increased calcium-to-sulfur ratio is greatly reduced, because of the relatively small quantities of sorbent involved. Uncertainties associated with sorbent selection and cost are also less significant.

Projections of particulate control and emissions of oxides of nitrogen for FBC power plants indicate that the more stringent emission requirements considered here of 0.03 lb/MBtu (12.9 ng/J) and 0.6 lb/MBtu (285 ng/J), respectively, are economically feasible and of lower cost impact than the more stringent sulfur oxide requirement. Conventional fabric filter (baghouse) techniques should permit achievement of this requirement, depending on particle size and future environmental standards. PFBC plants are projected to require two stages of particulate control equipment operating at the combustor temperature and pressure: e.g., conventional cyclones followed by a filter system. Nitrogen oxide levels from the assessment of FBC experimental results have been shown to be generally less than 0.6 lb/MBtu (258 ng/J) without special control efforts.

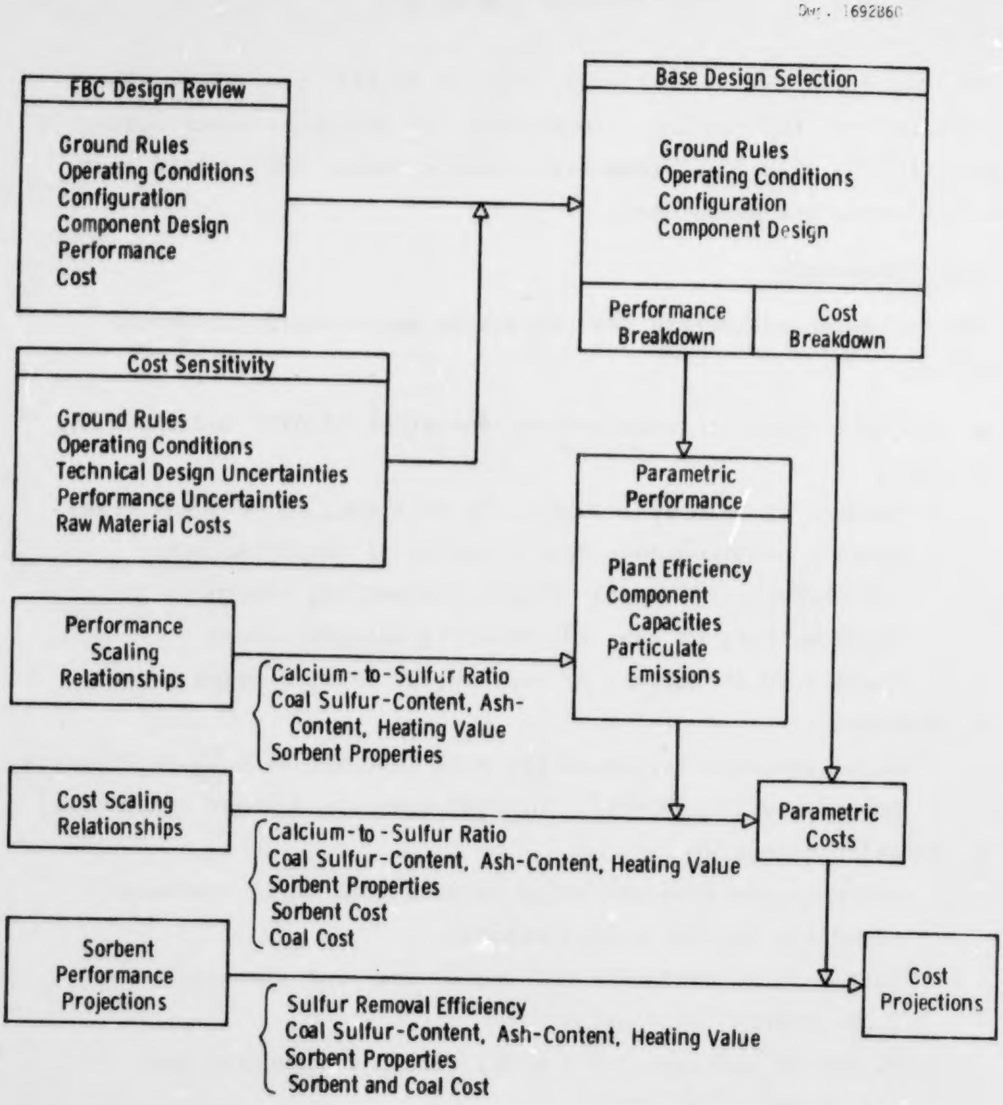
On the basis of available information, the projections developed indicate that both AFBC and PFBC should be able to achieve the higher levels of control considered in this evaluation economically if proper selection of combustor design and operating conditions is made. Development programs should focus attention on developing large-scale information on the relationship between combustor operating conditions and FBC plant emissions, while engineering evaluation of this information should be performed to provide assessment of FBC pollution control capabilities.

III. METHODOLOGY OF EVALUATION

The evaluation has utilized the results of several previous in-depth engineering studies to develop quantitative cost and performance projections. An evaluation logic-diagram (Figure 1) summarizes the sequence of steps. Previous FBC engineering studies were reviewed, and base designs (for AFBC and PFBC) and costing ground rules were selected. Relationships for scaling plant performance and plant cost were developed and applied to the base designs in order to generate plant performance and cost results with respect to the study parameters (calcium-to-sulfur ratio, coal properties, sorbent properties, coal cost, sorbent cost). Projections of sorbent performance and sulfur removal efficiency versus calcium-to-sulfur ratio, as a function of the combustor design and operating conditions, were applied to generate specific cost estimates for FBC power plants.

These costs were then compared with costs for conventional boilers with scrubbers, employing cost estimates generated in previous studies.[8,9]

Results from the evaluation are reported in three sections covering each of the potential revised emission standards categories: sulfur oxide control, particulate control, and control of oxides of nitrogen. The report body is followed by appendices that detail the sulfur oxide removal data base and model, the FBC power plant design bases and cost projections, particulate control projections, and the conventional power plant design basis and cost projections.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Sulfur Removal Efficiency
Coal Sulfur-Content, Ash-Content, Heating Value
Sorbent Properties
Sorbent and Coal Cost
Cost Projections

Figure 1 - FBC cost-evaluation logic diagram

IV. BASIS OF EVALUATION

The basis applied in this evaluation is briefly described. The study parameters, the costing assumptions, the FBC power plant base-designs, the FBC plant performance-projection basis, and the FBC cost-projection basis are summarized.

FBC Study Parameters

The following parameters and conditions were considered in the evaluation:

- Concepts - once-through sorbent operation of AFBC and PFBC
- Coals
  - Eastern (2 to 5 wt % sulfur, 10 wt % ash, 13 wt % moisture)
  - Western subbituminous coal, capable of complying with 1.2 lb/MBtu (516 ng/J) without further SO_2 control (0.6% sulfur, 8% ash, 23,260 kJ/kg heating value)
  - Lignite (0.4% sulfur, 6% ash, 18,600 kJ/kg heating value)
- Sorbents
  - Representative limestone for AFBC (average sorbent reactivity)
  - Representative dolomite for PFBC (average sorbent reactivity)
- Emission Standards
  - Sulfur oxide standard - 1.2 lb/MBtu (516 ng/J) (current standard) and 90% sulfur removal
  - Particulate standard - 0.1 (43.0) (current standard) and 0.03 lb/MBtu (12.9 ng/J)
  - Oxides of nitrogen - 0.7 (301) (current standard) and 0.6 lb/MBtu (258 ng/J)

- Operating Conditions
  - Calcium-to-sulfur ratio - 1 to 10
  - Fluid-bed combustor - 2 sets of conditions (temperature, pressure, velocity, bed depth, excess air, sorbent particle diameter) illustrated for each of the AFBC and PFBC power plants
- Sorbent Cost - \$5-60/Mg (\$5-66/short ton)
- Coal cost - \$1/MBtu (\$0.95/GJ).

The effect of coal price, coal ash-content, and heating value were also examined and found to be of secondary importance with respect to the relative economics of AFBC, PFBC, and conventional power plants.

Costing Assumptions

Table 1 summarizes the costing assumptions applied. These costing assumptions were adapted from the recent Energy Conversion Alternative Study (ECAS) ground rules, and were applied identically for both AFBC and PFBC.[8,10,11]

Power Plant Base Designs

Previous FBC power plant designs for AFBC and PFBC were reviewed.[1-3,8,9,11,12] Plant design comparisons are summarized in Appendix B. Base power plant designs were adapted from these studies, drawing heavily from the Westinghouse and General Electric plant designs resulting from the ECAS programs.[8,11] The General Electric designs for AFBC and PFBC power plants were applied with several modifications because they represent a consistent basis on which a conventional power plant design was also developed.[8] Details of the selected design bases are discussed in Appendix B.

The major FBC power plant design characteristics selected for the base plant were:

- Combustor configuration - modular design with vertical stacking of beds and separate steam generation functions for each bed; horizontal steam tubing and waterwalled vessels.

- Control of carbon utilization - carbon burnup cell (1100°C, 30% excess air)
- Coal and sorbent feeding - in-bed feeding with coal and sorbent premixed
- Particulate control equipment - final stage cleaning by baghouse (AFBC) and granular bed filter (PFBC)
- Spent solids handling - air cooling of spent solids with recovered energy used for coal and sorbent drying
- Spent solids storage - 15 day on-site storage with final, long-term, off-site storage in lined containment
- Steam cycle - 24,000 kPa, 538°C superheat temperature, and 538°C reheat temperature
- Gas turbine compression ratio (PFBC) - 10:1
- Cooling tower - wet mechanical draft
- Base plant capacity - the base plants had electrical capacities of about 800 MWe for AFBC and 900 MWe for PFBC, depending mainly upon the sorbent calcium-to-sulfur ratio and coal properties. Coal energy input was fixed at 2.262 GJ/s for the AFBC power plant and 2.308 GJ/s for the PFBC power plant to comply with the ECAS study basis.[8] These energy inputs remained fixed as coal characteristics and operating conditions were varied in the evaluation.

A multitude of potentially acceptable plant design options exist, as is apparent from Appendix B. Various degrees of modularity can be applied in an attempt to maximize shop fabrication of the combustors. Numerous combustor configurations are possible, ranging from horizontal arrangements of beds placed side-by-side to the vertical stacking concepts. Alternative steam circuitry and tubing layouts have been conceived.

The utilization of carbon can be maintained at high levels by several schemes: separate bed for combustion of elutriated char fines, operated at higher excess air and temperature (carbon burnup cell); high excess air operation of the combustor beds (up to 100% excess air

Table 1

COSTING ASSUMPTIONS

Site	- midwest
Cost base year	- mid-1975
Construction start	- mid-1975
Construction time, yr	- 5.5
Escalation, %/yr	- 6.5
Interest during construction, %/yr	- 10.0
A&E services, % of indirect cost	- 15.0
Labor rate, $/hr	- 11.75
Indirect field labor cost, % of direct field labor	- 90
Contingency, % of direct, indirect, and A&E services	- 20
Plant availability target, %	- 90
Plant capacity factor, %	- 65
Capitalization rate, %/yr	- 18
Operating and maintenance cost, %/yr	- 1.95
Coal price, $/GJ	- 0.95

for a pressurized fluid-bed boiler); recycle of elutriated char fines back to the combustor beds; and minimization of char elutriation by low-velocity operation, higher bed temperatures, control of coal particle size, etc.

The feeding of coal and sorbent to the combustor may be performed either separately or in combination, with feeding being in bed or above bed. Several techniques for distributing the solids to a multitude of feed points and injecting the solids are available.

Final-stage particulate control can be achieved by several methods. For AFBC, high-efficiency cyclones, electrostatic precipitators (high or low temperature), or fabric filters may be applied. For PFBC, high-efficiency cyclones, a hot electrostatic precipitator, granular bed filters, ceramic filters, or low-temperature cleaning schemes combined with a regenerative heat exchanger may be applicable.

Spent solids from FBC power plants may find specific forms of utilization or alternative disposal applications. Long-term off-site storage may be required in some cases, depending on plant location.

Thus, numerous FBC power plant design and operating options have been considered for both AFBC and PFBC plants utilizing the basic design concept selected.* The cost data for these options indicate that the FBC power plant equipment investment will not be greatly sensitive to the choice of these options, but the performance and operability of the plant could be significantly influenced by their choice (see Appendix B). The options selected for this evaluation provide reasonable plant cost projections and assume that any significant performance and operability problems have been resolved.

Performance Projection Basis

The FBC power plant performance factors (equipment capacities, plant efficiency, sulfur removal, etc.) that were judged to be the most signifi-

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*Alternative FBC concepts are being considered that may have an appreciable impact on system cost (e.g., recirculating bed concepts[2]). This report does not include an analysis of the impact of emission standards on these systems.

Table 2

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APPROACH FOR CONSIDERATION OF FBC PERFORMANCE VARIABLES

Scaling Parameters — calcium-to-sulfur ratio, coal properties (sulfur content, ash content, heating value), sorbent properties, and combustion conditions

Equipment Capacities Scaled (Appendix C) —

- Sorbent handling, crushing, drying and feeding
- Fines handling
- Spent sorbent cooling, handling, and on-site storage
- Spent solids off-site disposal

Power Plant Thermal Efficiency — contributing factors scaled (Appendix C)

- Auxiliary power consumption
- Combustion efficiency
- Energy losses — sorbent calcination, sensible heat losses, etc.

Sulfur Removal Performance (Appendix A)

- Extensive kinetic data collected on TG equipment, and comprehensive experimental FBC data bank utilized
- Sulfur removal model utilized to project FBC sulfur removal efficiency as a function of: 3
  - Calcium-to-sulfur ratio
  - Bed depth
  - Velocity
  - Sorbent particle size
  - Bed voidage and density
  - Sorbent calcination conditions
- Results of model compare well with experimental FBC data from data bank for broad range of operating conditions and large number of data sources, but data from large combustors and data for high levels of SO_2 removal are limited.

Particulate Control Projections (Appendix D)

- Experimental fresh feed size distribution selected
- Coal-ash elutriation rate projected as 100% of coal ash content
- Sorbent attrition rate and elutriation rate projected from basic attrition model 13
- Experimental size distribution of elutriated coal-ash and sorbent fines applied, as available
- Experimental grade efficiencies for cyclones, electrostatic precipitators, baghouses, and granular bed filters applied
- Computer model projects particulate emission from FBC particulate control system

NO_x Control Projections

- Statistical evaluation of FBC NO_x emission data performed
- Projections based on representative NO_x performance without specific control efforts

cant in influencing the power plant cost were scaled from the base plant performance values (from General Electric base plants[8]) with respect to the study parameters. The nature of the performance projections are summarized in Table 2. The relation developed and applied in this evaluation to estimate the effect of these variables on plant performance are simplified to provide practical insight and sufficient accuracy for plant cost projections. These simplified relations are detailed in Appendix C.

COST PROJECTION BASIS

Investment and energy costs were estimated by scaling the base plant equipment costs (from the modified General Electric base plant[8]) and plant performance (electric generating efficiency), derived in Appendix B, with respect to the calcium-to-sulfur ratio and the coal properties. This procedure for scaling is detailed in Appendix C.

For the FBC cost projections, seven plant categories were used to describe the FBC capital investment breakdown.[8,11] Five of these categories were assumed to be fixed in cost over the ranges of parameters considered (turbine generators, electrical, civil and structural, process piping and instrumentation, and yardwork and miscellaneous), and two categories were scaled according to changes in capacity (steam generators and process mechanical equipment).

Component costs, direct labor costs, indirect field costs, and materials costs were scaled according to capacity ratios with a 0.85 power factor generally appreciable for solids handling items, a 0.6 power factor for cyclones and particulate control auxiliary components, and a 0.68 power factor for fans and motors.

The process mechanical equipment category includes the scaled items coal drying and crushing, limestone drying and crushing, coal and limestone blending and feeding, spent bed material cooling, and spent solids material on-site and off-site storage. Solids handling controls are assumed to be constant in cost. Complete listings are presented in Appendix C.

The steam generator category includes hot gas cleanup items, air supply, and the combustor and steam generator items. The hot gas cleanup items, which are assumed to be unchanged in cost with respect to the study parameters, are the bed cyclone units, cyclone air lock valves, fines injection system, carbon burnup cyclone air lock valves, and cooler air lock valves, these items being generally more strongly influenced by gas-handling capacity than by the variables considered in this study. A baghouse (AFBC) has been substituted for the base design ESP. A granular bed filter (PFBC) based on previous Westinghouse designs has been used in place of the General Electric design. The fluid-bed combustor/steam generator is assumed to be constant in cost for the parameters of interest in this study, based on previous sensitivity analyses which are reviewed in Appendix B. The maximum increase in combustor cost resulting from reduced gas velocities would be about 10 percent expressed as $/ton of coal but may be reduced when expressed as $/kw if the plant performance improves with reduced velocity.

The cost of electricity is obtained from the total plant capital investment and operating cost, and the plant generating capacity. The categories considered are capitalization, operating labor, indirect operating costs and maintenance costs, the cost of coal and the cost of sorbent.

COST SENSITIVITY AND UNCERTAINTIES

The sensitivity of the FBC power plant investment to the design and operating conditions of the fluidized-bed combustion system (not including solids feeding and particulate control) has been shown to be rather small (no more than ±10%).[2] This is partly because the FBC power plant is largely made up of conventional power plant components (see Appendix B). The solids feeding system, the fluid-bed combustor, and the hot gas cleaning system represent the developmental components which account for 20 to 30 percent of the total plant cost. The fluid-bed combustor itself accounts for 10 to 15 percent of the plant investment and is very insensitive to changes in design and operating conditions.[2] The combustor should be designed to maximize performance (e.g., carbon utilization, sulfur removal efficiency, particle carry-over, etc.) rather than to minimize the vessel cost. For example, low combustor fluidization velocities will probably result in a power plant producing cheaper electricity (due to improved process performance) than does a FBC power plant designed with a compact fluid-bed combustor (high fluidization velocities and large sorbent particles).

The greatest technical design uncertainties appear to be associated with the solids feeding and hot gas cleaning systems, especially for PFBC. These systems could have a large impact on plant reliability.

The greatest performance uncertainty is undoubtedly associated with the rate of sorbent feeding required to achieve the specified sulfur removal efficiency. Significant variability in sorbent properties and performance exists. Sorbent performance data from large FBC units, and for high degrees of SO_2 removal, are lacking.

The costs of coal and sorbent are subject to great uncertainty and variability, depending on power plant location. These uncertainties could have a greater impact on the plant energy cost than do technical design uncertainties or performance uncertainties.

Finally, the cost assumptions, or ground rules, applied (Table 1) may provide substantial cost uncertainties. Plant site is very influencial in this respect. Power plant construction time could vary significantly from one plant to another. Even though plant costs (AFBC, PFBC, and conventional) are compared, with the same ground rules applied to all of them, impacts of differing site conditions could be significantly different for each of them. The concept of "economy of scale" is also very uncertain because of the impact of plant construction time, interest during construction and escalation, and regulatory restrictions, and different plant capacities may modify the comparison of costs.

Within the framework of the set of costing assumptions stated, the FBC power plant costs projected are expected to be broadly representative of FBC power plant cost potential.

V. SULFUR-OXIDE CONTROL

The control of sulfur oxide emissions from the FBC power plant is addressed independently of particulate control and oxides of nitrogen in this section. Plant performance and cost is expressed parametrically as a function of calcium-to-sulfur ratio and coal properties. The impact of combustor operating conditions on the required calcium-to-sulfur ratio is discussed and related to specific projections of performance and plant cost. Costs are compared with a conventional coal-fired power plant using lime slurry scrubbing for sulfur oxide control, as a function of sorbent cost.

Two sulfur oxide emission requirements are considered, the current standard of 1.2 lb SO_2/MBtu (516 ng/J) and a more stringent requirement of 90 percent sulfur removal. The relationship between the sulfur oxide standard expressed in lb SO_2/MBtu and expressed in percent removal is shown in Figure 2. For a 4.0 wt % sulfur Eastern bituminous coal (10 wt % ash, 13% moisture, heating value of 25,300 J/g) the required sulfur removal efficiency to satisfy the current standard is about 83 percent.

POWER PLANT OVERALL ENERGY CONVERSION EFFICIENCY

The rate of sorbent feeding to the FBC power plant can have a significant impact on the power plant thermal efficiency, due mainly to the energy requirement for calcination of the sorbent. Increased auxiliary power consumption for solids handling is of secondary importance. Plant efficiency is expressed as a function of calcium-to-sulfur ratio and coal sulfur content in Figure 3, considering efficiency losses due to calcination and auxiliary power consumption. Both PFBC and AFBC are considered. The figure assumes a constant coal combustion efficiency and a constant rate of heat release due to the sulfation reaction for all calcium-to-sulfur ratios equivalent to meeting the current sulfur oxide standard for each coal sulfur content. The PFBC plant efficiency drops more

Curve 693150-A

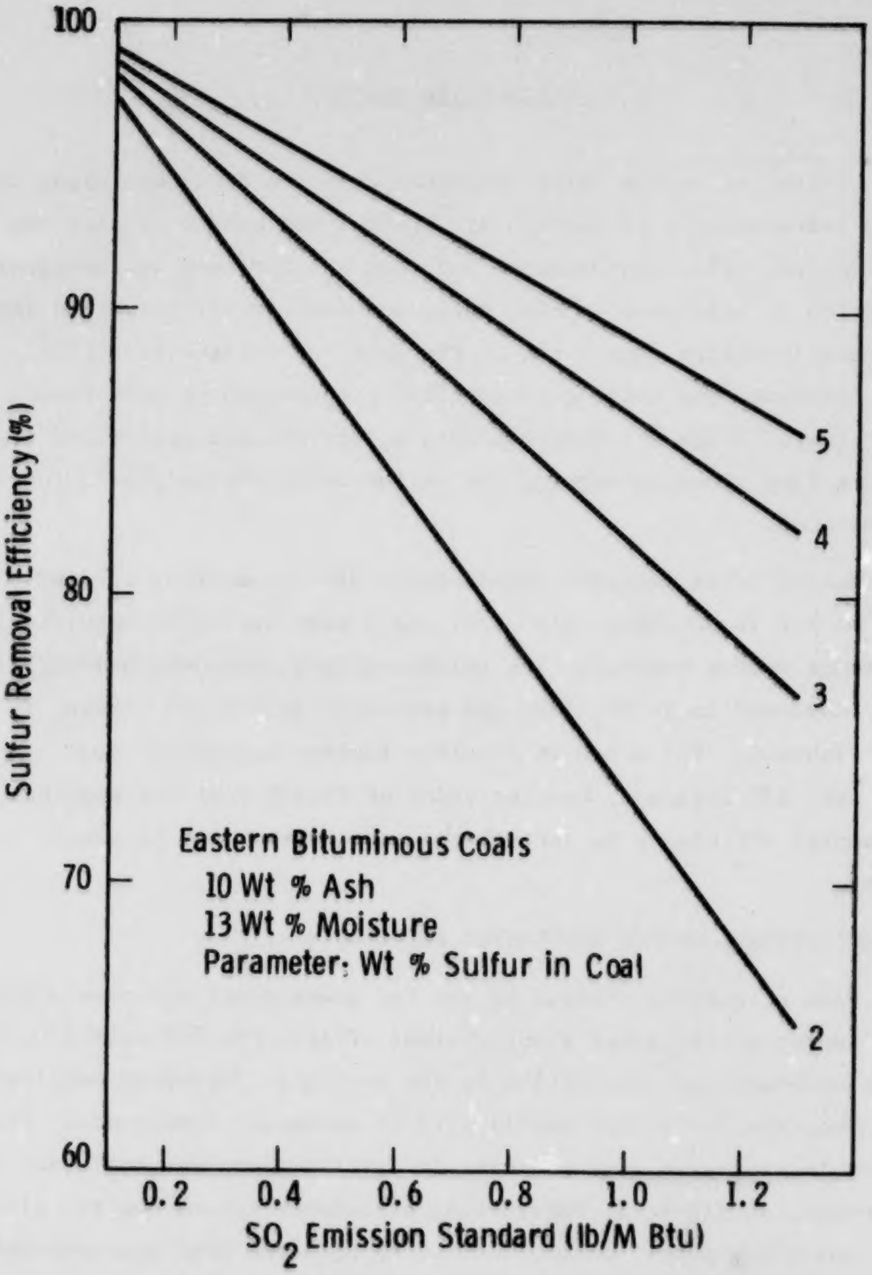


Figure 2 - Required sulfur removal efficiency

Curve 693149-A

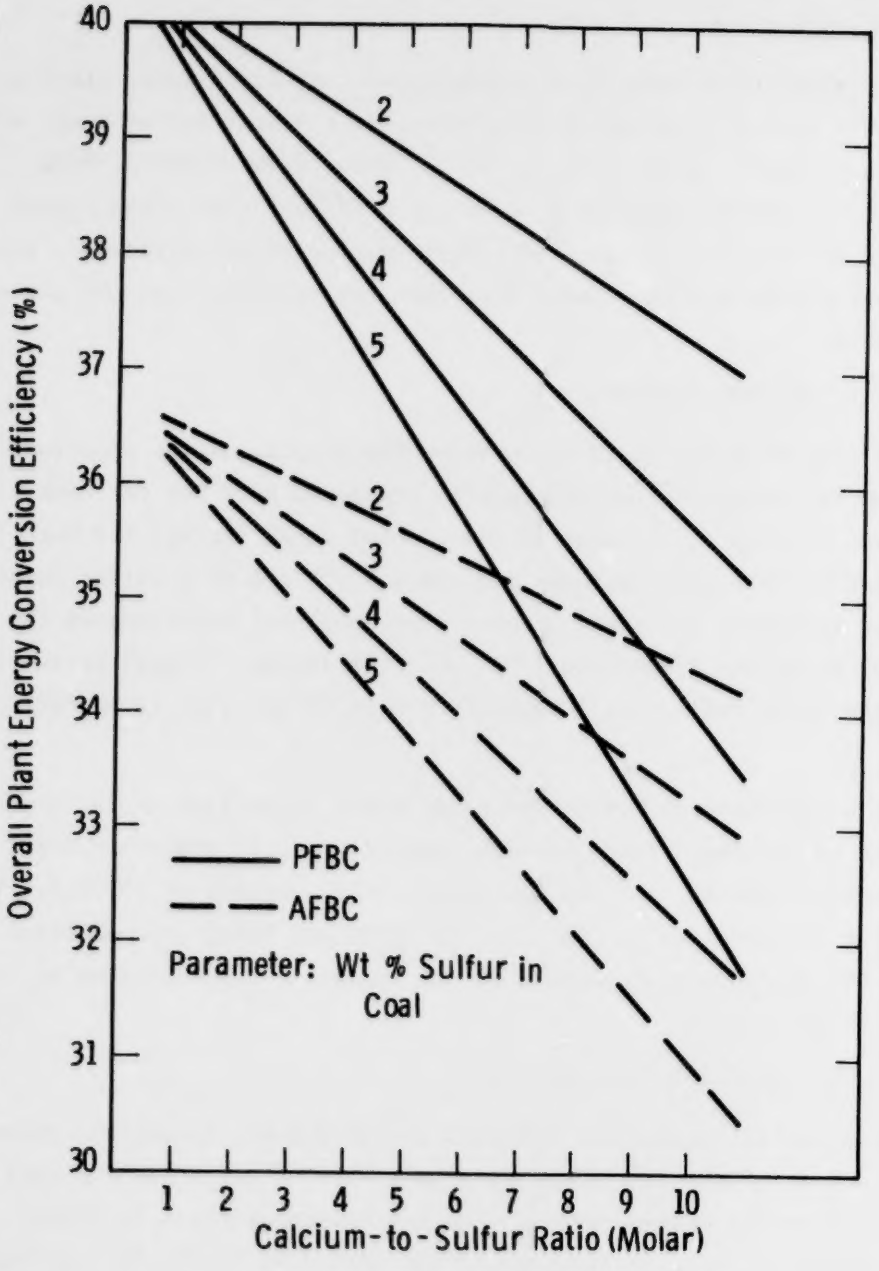


Figure 3 - FBC power plant energy conversion efficiency

swiftly than does the AFBC plant efficiency because the dolomite sorbent for PFBC has a greater calcination energy loss than does the limestone sorbent used for AFBC.

The energy efficiency of a conventional coal-fired power plant with lime slurry scrubbing is about 34 percent for a 4 wt % sulfur coal, with a stack-gas reheat temperature of 79°C.[8] Then, from Figure 3, PFBC power plants should always be more energy efficient than conventional power plants over the range of calcium-to-sulfur ratio considered, and AFBC power plants should be more efficient for calcium-to-sulfur ratios less than 6.

POWER PLANT RESIDUE GENERATION

The rate of spent solids generation for disposal is an important environmental factor for which specific standards have not yet been developed. The ratio of spent solids to coal input (kg/kg) for both AFBC and PFBC will range between 0.3 and 0.7 for a 4 wt % sulfur Eastern coal (not including the ash content of the coal) and depending on the combustor operating conditions. The ratio of (spent sorbent) to the coal ash in the solid residue will range from 2 to 10 for typical Eastern coals.

For a conventional coal-fired power plant using lime slurry scrubbing the ratio of scrubber sludge to coal input (kg/kg) is about 0.4 for a 4 wt % sulfur Eastern coal and assuming a water content of 50 wt %. This is a value intermediate to the range for AFBC and PFBC. A comparison between the environmental impacts of the two solid waste sources is being developed.

POWER PLANT INVESTMENT AND ENERGY COST

Total capital investment for PFBC and AFBC power plants are shown in Figure 4 as a function of the calcium-to-sulfur ratio and the coal sulfur content for Eastern coals. The derivation of the base plant capital costs is shown in Appendix B; the method by which the base plant costs were varied to account for variations in calcium-to-sulfur ratio is detailed in Appendix C. The base particulate control equipment is assumed constant for all calcium-to-sulfur ratios (baghouse for AFBC and granular bed filter for PFBC). The increasing investment with increasing

Curve 693146-A

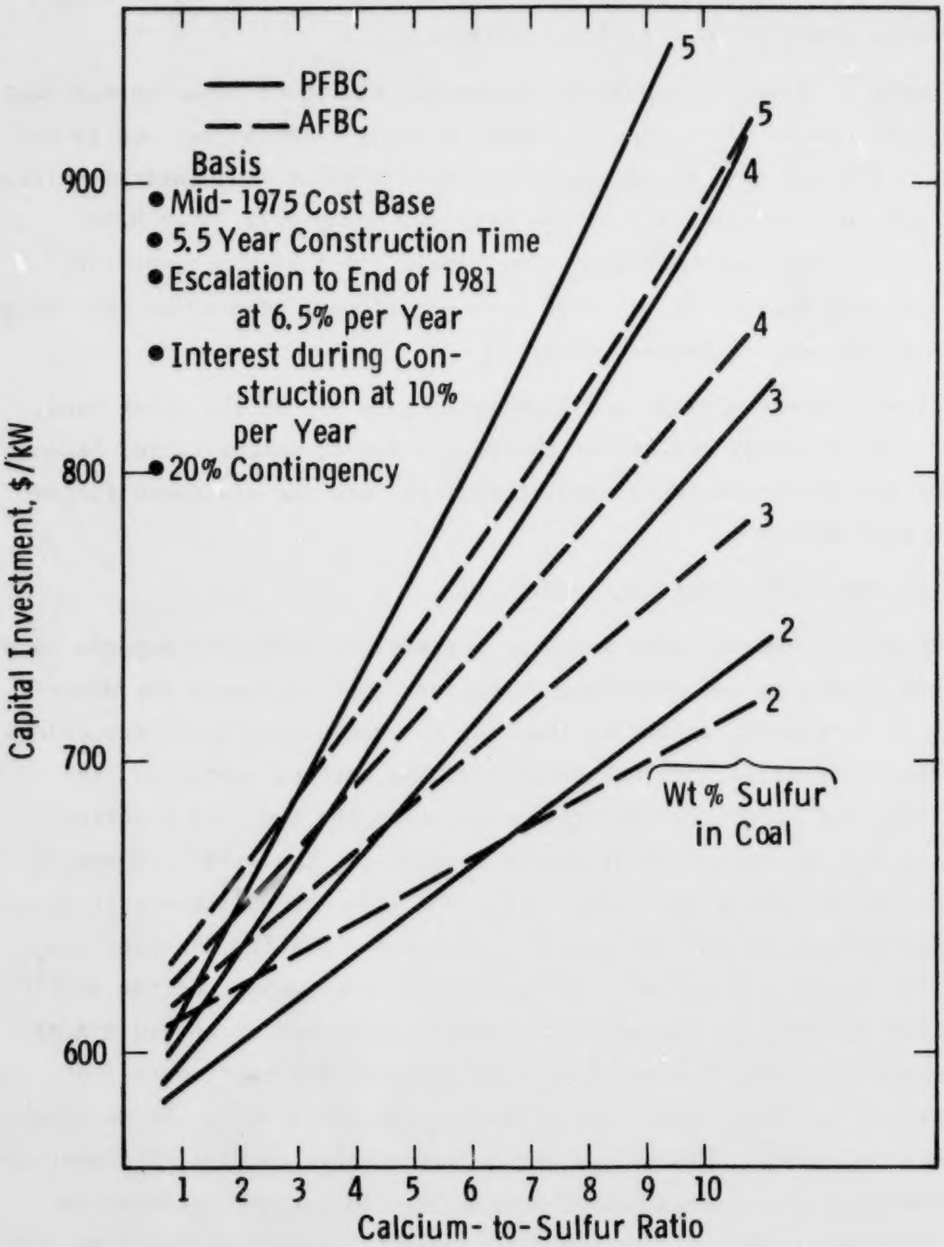


Figure 4 - Capital investment for FBC with Eastern coals

calcium-to-sulfur ratio is due to the combined effects of increased equipment cost for solids handling, crushing, drying, feeding and disposal, and reduced power plant efficiency (Figure 3).

Figure 5 shows the resulting parametric FBC power plant energy cost for Eastern coals. A sorbent delivered cost of $5/Mg is assumed in the figure. PFBC and AFBC energy costs are very similar at identical values of the calcium-to-sulfur ratio. Higher coal costs will favor PFBC because of higher plant efficiencies. Again, the calcium-to-sulfur ratio and coal sulfur content have very significant impacts on the energy cost with high-sulfur Eastern coals.

FBC with Western coals and lignite (Figure 6), on the other hand, are not significantly influenced by the calcium-to-sulfur ratio, because of their low sulfur contents, and, therefore, are not discussed further in this evaluation.

IMPACT OF COMBUSTOR OPERATING CONDITIONS

The sulfur removal efficiency of a fluid-bed combustor depends on a multitude of design and operating conditions that influence the distribution and release of sulfur in the bed, the contacting of sulfur oxides and sorbent particles, the calcination of the sorbent particles, the temperature and concentration distributions in the bed, the reaction kinetics, the interparticle diffusion resistance, the attrition and elutriation of sorbent particles, etc. The major operating conditions have been identified and correlated (for dense-phase fluidization systems) with sulfur removal efficiency (see the development of the Westinghosue kinetic model in Appendix E of Reference 4, and in Appendix A of this report). Figure 7 shows performance curves for representative sorbents and specific operating conditions projected using the Westinghouse kinetic model. The curves are representative of typical limestones and dolomites, though significant variability in sorbent performance exists (see Appendix A). The specific sorbent types that served as the basis for Figure 7 (Carbon limestone and 1337 dolomite) are among the more reactive of the sorbents that have been tested on the Westinghouse

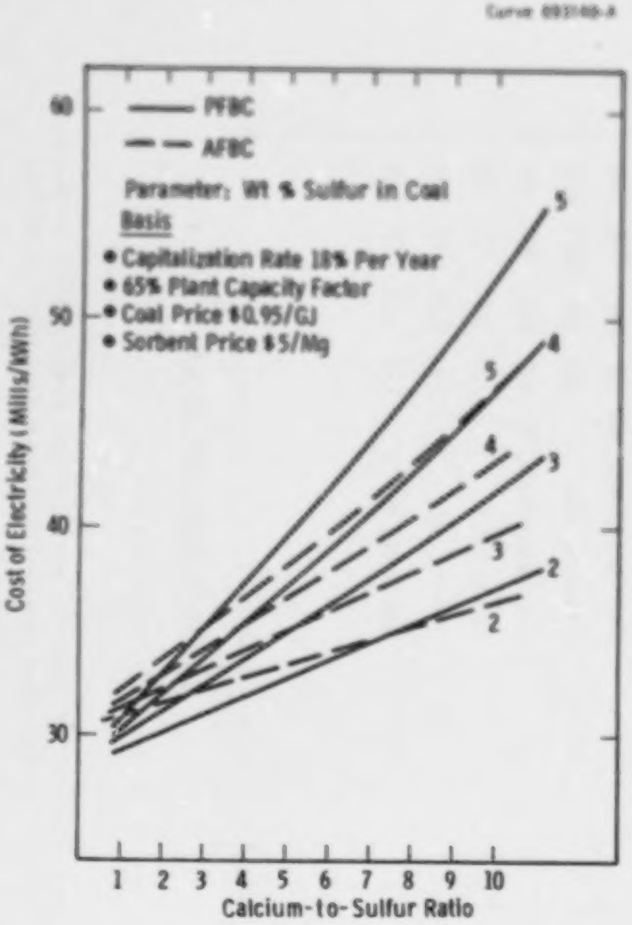


Figure 5 - Cost of electricity for FBC with Eastern coals

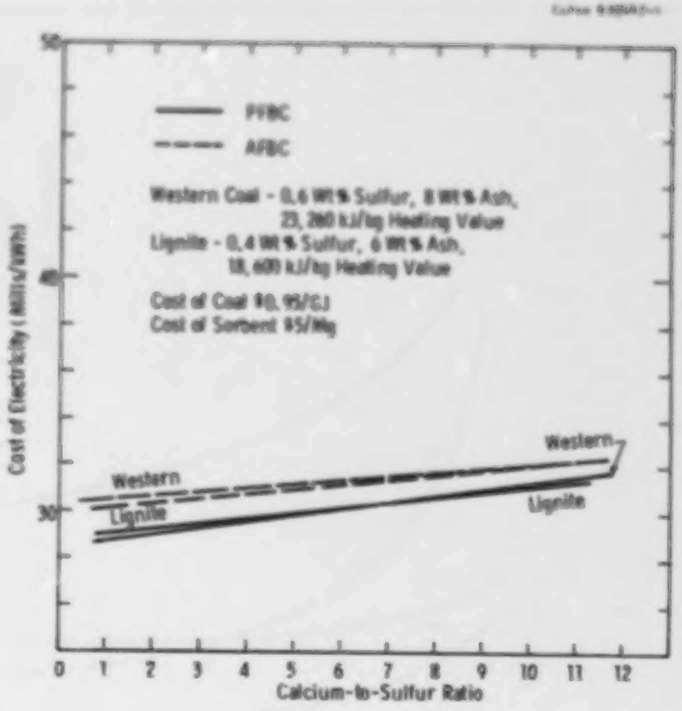
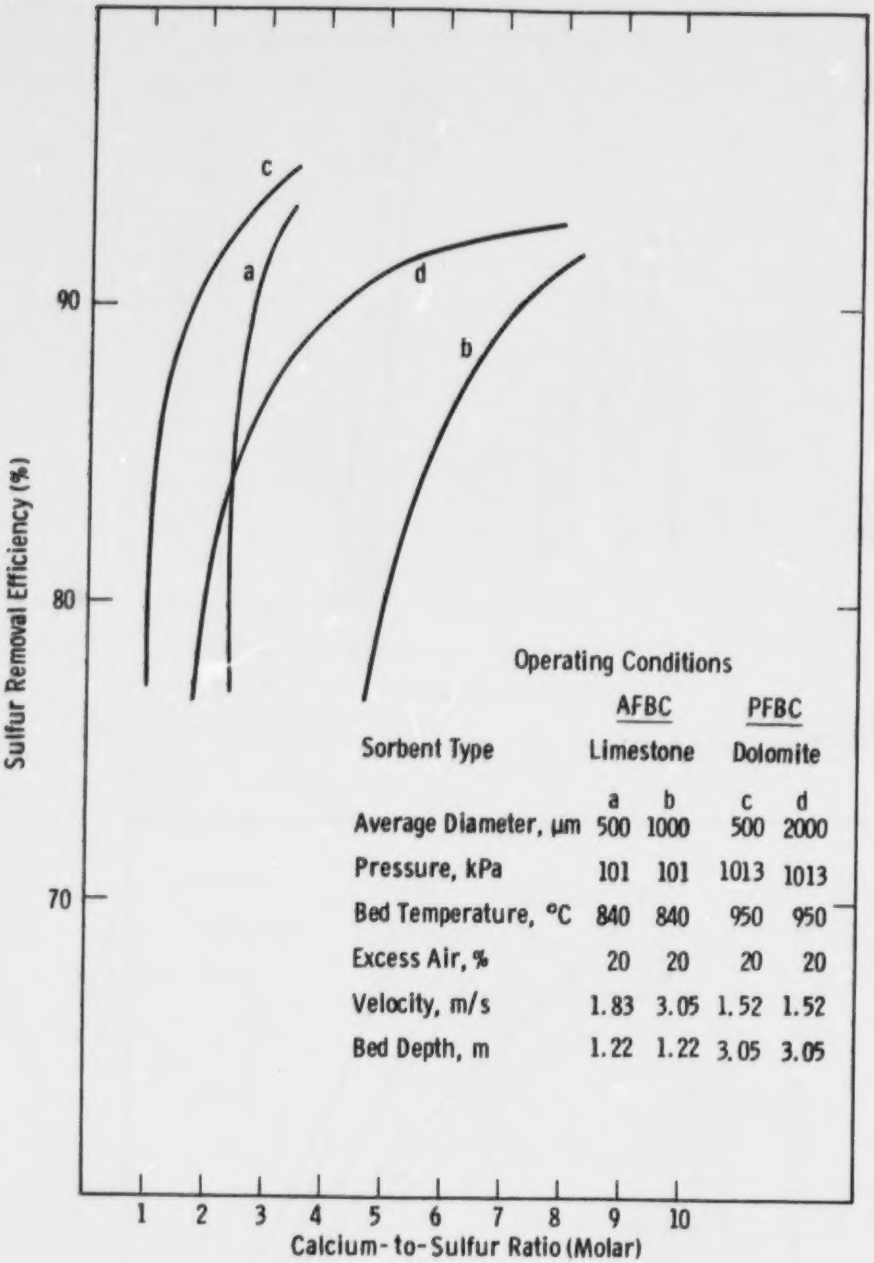


Figure 6 - Cost of electricity for FBC with Western coal and lignite

Curve 693151-B



Operating Conditions

	AFBC		PFBC	
Sorbent Type	Limestone		Dolomite	
	a	b	c	d
Average Diameter, μm	500	1000	500	2000
Pressure, kPa	101	101	1013	1013
Bed Temperature, °C	840	840	950	950
Excess Air, %	20	20	20	20
Velocity, m/s	1.83	3.05	1.52	1.52
Bed Depth, m	1.22	1.22	3.05	3.05

Figure 7 - Sulfur removal performance for typical sorbents (projected using Westinghouse kinetics model)

thermogravimetric apparatus and are expected to be representative of sorbents than will be available to FBC power plants. The performance projections in Figure 7 require confirmation on operating fluidized-bed boilers sufficiently large to provide data representative of commercial-scale units.

Table 3 lists two sets of combustor operating conditions and projected calcium-to-sulfur ratios required to meet the two levels of sulfur oxide emissions (1.2 lb/MBtu [516 ng/J] and 90% sulfur removal efficiency) for a 4 wt % sulfur Eastern coal. The calcium-to-sulfur ratios were taken from Figure 7. The sorbent average particle diameter and fluidization velocity (gas residence time in the bed) are the key conditions varied between the two sets, and they clearly have a dramatic impact on the required calcium-to-sulfur ratio. The high velocity-large particle sorbent conditions are typical of many designs that have been proposed.

With these sets of operating conditions to illustrate the impact of the operating conditions selection and the potential performance of FBC power plants, a comparison may be made between the energy cost of AFBC, PFBC, and conventional coal-fired power plants. None of the sets of conditions listed in Table 3 are meant to represent optimum operating conditions.

The energy cost projections for the conventional coal-fired power plant with lime slurry scrubbing have been taken from two sources. The cost for the current sulfur oxide emission of 1.2 lb/MBtu (516 ng/J) was extracted from a recent report by TVA which applied the same cost ground rules as were applied in this report.[9] The cost agrees closely with those from previous TVA cost studies for FGD systems when placed on the same basis.[14] The cost for 90 percent sulfur removal was taken from the General Electric ECAS cost by applying cost modifications suggested by TVA. The details of these projections are presented in Appendix E.

Energy costs for PFBC, AFBC, and a conventional plant with lime slurry scrubbing are plotted together in Figures 8 and 9 as a function

Table 3

FBC PLANT OPERATING CONDITIONS

	AFBC		PFBC	
Pressure, kPa	101		1013	
Bed temperature, °C	840		950	
Excess air, %	20		20	
Sorbent type	Limestone		Dolomite	
Sorbent diameter, μm	1000	500	2000	500
Bed velocity, m/s	3.05	1.83	1.52	1.52
Bed depth, m	1.22	1.22	3.05	3.05
Gas residence time in bed, s	0.4	0.67	2.0	2.0
Ca/S for 1.2 lb/MBtu SO_2 (516 ng/J) emission[a,b]	5.5	2.4	2.0	1.2
Ca/S for 90% sulfur removal efficiency[b]	7.0	2.9	4.5	1.7

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a - 83% sulfur removal for 4 wt % sulfur coal.

b - From Figure 7, projections based upon Westinghouse kinetic model. For sorbents of representative reactivity, based upon about 25 sources of sorbents that have been tested by Westinghouse on their laboratory thermogravimetric analysis apparatus in several hundred characterization tests (see Appendix A).

of the sorbent cost for the two sulfur oxide emission standards: 1.2 lb/MBtu (516 ng/J), the current standard, and 90 percent sulfur removal efficiency (equivalent to 0.74 lb/MBtu [318 ng/J]), both for a 4 wt % sulfur Eastern coal. Calcium-to-sulfur ratios in FBC are taken from the projections in Table 3.

For the current sulfur oxide standard, the break-even cost of sorbent ranges between \$7 and \$65/Mg for AFBC and is \$57 to >\$100/Mg for PFBC, depending upon the selected operating conditions. For 90 percent sulfur removal, the break-even cost of sorbent ranges between \$3 and \$55/Mg for AFBC and \$11 to \$70/Mg for PFBC, depending on the selected operating conditions. Typical sorbent costs in this country today range from \$5 to \$10 per Mg (at quarry), suggesting that FBC should be competitive with FGD.

The equivalent cost of spent solids disposal (for off-site disposal) that is included in Figures 8 and 9 is \$3/Mg of fresh sorbent for AFBC and \$4/Mg of fresh sorbent for PFBC. Potential markets for the spent solids would have some impact on the comparison in Figures 8 and 9 but, more importantly, could significantly reduce the environmental impact of the FBC spent solids.

A definitive analysis of the environmental impact of the disposal of spent sorbent from FBC processes cannot be carried out since environmental impact criteria are not established for land disposal. Data from leaching tests, however, provide a preliminary comparison of the environmental impact of fluidized-bed combustion and limestone scrubbing residues. Results are presented in Table 4 which indicate the FBC spent sorbent may have a reduced impact from leaching.[15] Codisposal of the FBC spent sorbent and ash is recommended since it may further reduce the environmental impact. The FBC spent sorbent and ash will form solid compacts with the addition of water, which results in an effective low-cost method to reduce the permeability and thus the environmental impact.[7,15]

The cost for disposal will depend on the alternative sorbent selected and the specific location. TVA[9] projected costs for AFBC, PFBC, and limestone scrubber residues, providing perspective on the disposal costs.

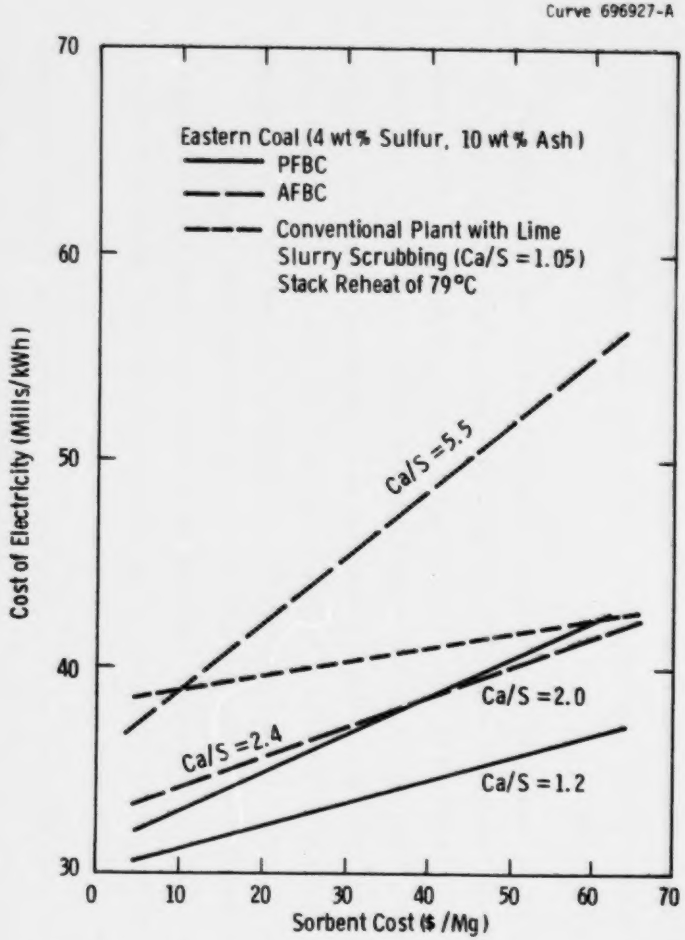

Figure 8 - Comparison between AFBC, PFBC and FGD for SO_2 emission standard of 1.2 lb/MBtu (516 ng/J) (current standard)

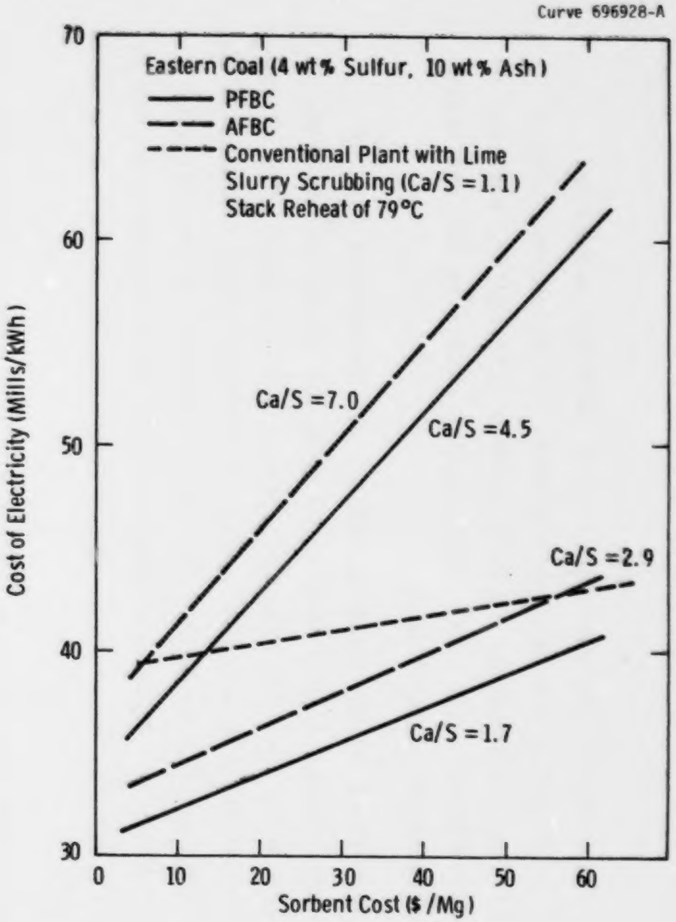

Figure 9 - Comparison between AFBC, PFBC and FGD for SO_2 emission requirement of 90% sulfur removal

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TABLE 4
PRELIMINARY COMPARISON OF THE ENVIRONMENTAL IMPACT FROM THE LEACHING OF FBC* AND FGD RESIDUES

	FBC	FGD
Chemical Property	① Solubility of Major Compounds: Ca, SO_4 and TDS Contributing to Potential Environmental Concern	① High Concentrations of Mg. Cl in Addition to Ca, SO_4, and TDS. (Plus Na in the Case of Double-Alkali System)
	② Major Component $CaSO_4$ Relatively Inert	② Major Component $CaSO_3 \cdot 1/2\ H_2O$: Potential Environmental Hazard for Untreated Sludge
	③ High Alkalinity in Leachate: pH <0 to 13	③ pH = 5 to 10 for Lime or Limestone Scrubbing Systems pH = 12 to 13 for Double-Alkali System
	④ Trace Elements: Not Expected to Cause Environmental Problem. Most Leachates Meet Drinking Water Standards	④ • Several Elements in Liquor and Leachate, e.g. As, Se, Cd, Mn NO_3, and F, Exceeding the Drinking Water Standards Including the Ponded and Oxidized Sludges • Improvement with Stabilization
	⑤ TOC in Leachate: Low	⑤ TOC in Leachate: Low TOC in Liquors: Low

* AFBC and PFBC

The comparative cost for on-site disposal based on fixation of the scrubber sludge with a fixing agent and formation of solid compacts with the FBC spent sorbent using water has been calculated on the basis of TVA report costs. The capital and operating cost result in a cost of about $7/ton ($8/Mg) dry residue for AFBC and PFBC plants and $12/ton ($13/Mg) sludge for limestone wet scrubbing plants. The quantity of spent sorbent from FBC processes will generally range from 0.3 to 0.7 kg/kg coal for a once-through sorbent utilization and will depend on the sorbent, coal, FBC design, and emission requirement. The comparative quantity of solids from the limestone wet scrubber will be approximately 0.4 kg sludge/kg coal or 0.2 kg dry solids/kg coal.

The impact of averaging time criteria related to the sulfur oxide emission standard has not been factored into this evaluation. Variations in coal sulfur content, sorbent properties, or combustor operability (coal and sorbent feeding, bed maldistribution, etc.) with time could reduce FBC sulfur removal reliability.

The effect of coal ash-content and coal heating value on the economics of FBC with high-sulfur Eastern coals is of secondary importance compared to the effect of coal sulfur-content and calcium-to-sulfur ratio. These secondary variables are not discussed in this report.

ADVANCED SULFUR REMOVAL SYSTEMS

More stringent sulfur oxide emission standards may lead to increased economic incentives to develop advanced sulfur removal systems for FBC. The small-scale investigation of several advanced concepts has been or is being performed. For example:

- Sorbent regeneration
    - Reductive decomposition
    - Two-step sulfide generation, H_2S generation
    - Sulfide-sulfates solid-solid reaction
- Precalcination of once-through sorbents for improved reactivity and sorbent utilization

- Additives for improved sorbent utilization
- Reconstitution of sorbent fines by agglomeration
- Alternative metal oxide sorbents.

All of these advanced concepts could potentially reduce the rate of sorbent consumption and improve FBC economics and environmental impact.

VI. PARTICULATE CONTROL

Particulate control in AFBC and in PFBC are quite different in nature. For AFBC conventional technology exists - hot electrostatic precipitators (~400°C) or fabric filter techniques (baghouses) - which should be capable of controlling particulate emissions to the 0.03 lb/MBtu (12.9 ng/J) level. Testing of these control options on operating fluidized-bed combustors, however, is necessary before actual control performance and any operating problems can be firmly defined. No such testing has yet been conducted.

For PFBC the protection of the gas turbine from erosion and deposition damage may require that particulate emissions be controlled to levels probably less than 0.03 lb/MBtu, depending on size distribution and composition. Granular bed filters, ceramic filters, and high-temperature precipitators have been proposed for hot gas cleaning for PFBC but have not yet been demonstrated.

The total loading and particle size distribution of particulates attrited and elutriated from a fluidized-bed combustor have been projected, using a Westinghouse model; these projections have been compared with the expected performance of particle control equipment that might reasonably be employed (see Appendix D). The projections suggest that, over the range of calcium-to-sulfur ratios considered (up to 10), within the variability of sorbent attrition performance expected, and based on the current data available, particulate emission levels of 0.03 lb/MBtu (12.9 ng/J) should be economically achievable for AFBC and PFBC, based upon the particulate control equipment costs included by Westinghouse in this study (see Appendix B).

For AFBC the major uncertainties involve the details of the equipment specifications (filter materials, face velocities, etc.) and operating and maintenance procedures. The uncertainties involved in the

PFBC particulate control system are greater, but the level of the emission standard may have little impact upon the development of these systems because of the turbine protection requirements, which are expected to be controlling unless the particulate size distribution is relatively fine. If future emission requirements are established that require control of fine particulates (<10 μm), then environmental concerns may become controlling.

As with sulfur removal control, the combustor design and operating conditions have a great impact on the control of particulate emissions. With little cost penalty the combustor can be designed and operated (e.g., at low velocities) to limit particle attrition and elutriation, though very efficient final-stage particulate control devices will still be required. The coal ash-content and coal heating value have little effect on the particulate control requirements for FBC relative to the coal sulfur-content and the calcium-to-sulfur ratio.

VII. CONTROL OF OXIDES OF NITROGEN

A statistical evaluation of FBC NO_x emissions data has been performed under contract to EPA (EPA contract number 68-02-2132). Modeling of the formation of oxides of nitrogen in fluid-bed combustors is also being attempted. The data from PFBC units show a high probability of emissions significantly lower than 0.6 lb NO_2/MBtu (258 ng/J). AFBC emissions should satisfy a 0.6 lb NO_2/MBtu (258 ng/J) requirement but are not generally as low as in PFBC.

For the control of sulfur oxides and particulates, the control techniques and relation to FBC operating and design parameters are relatively clear. For nitrogen oxides, however, the relation between emissions and operating and design parameters in FBC are not clearly understood. Efforts are under way to improve understanding of these relationships and to develop NO_x control strategies that can reduce FBC NO_x emissions below their inherently low levels.

VIII. CONCLUSIONS

- On the basis of available information, the more stringent emission requirements considered in this study (sulfur oxides: 90% sulfur removal; particulates: 0.03 lb/MBtu (12.9 ng/J); oxides of nitrogen: 0.6 lb NO_2/MBtu (258 ng/J) should be economically achievable for both AFBC and PFBC power plants.
- The economical realization of these environmental goals depends critically on the proper selection of fluid-bed combustor design and operating conditions. In particular, the gas residence time in the bed (as determined by gas velocity and bed height) should be sufficiently long, and sorbent particle size should be sufficiently small. In this assessment residence times of 0.67 to 2.0 s (gas velocities of 1.5 to 1.8 m/s) and partial size averaging 500 μm appeared to offer effective SO_2 removal performance, although these conditions are not necessarily optimal.
- The high level of sulfur oxides emission control considered has a greater impact on the FBC power plant energy cost than do the revised particulate and oxides of nitrogen standards considered. The most critical process parameter, with respect to FBC power plant cost and performance, is the calcium-to-sulfur ratio.
- The fluid-bed combustor cost is not strongly dependent on changes in design and operating conditions. The fluid-bed combustor should be designed to minimize plant energy cost rather than to minimize the combustor cost. For example, low - rather than high - fluidization velocities will probably result in lower FBC power plant energy cost.

- Particulate control to levels as low as 0.03 lb/MBtu (12.9 ng/J) should be economically achievable for AFBC using commercially available techniques. Baghouses seem most suitable for this duty. However, no testing of any type of final-stage particle control device on an AFBC unit has yet been conducted.
- Particulate control to levels below 0.03 lb/MBtu (12.9 ng/J) may be dictated for PFBC by turbine protection requirements, depending on particle size. Projections indicate that 0.03 lb/MBtu should be achievable, but the technology to meet this control at high temperature and pressure has not yet been demonstrated.
- Oxides of nitrogen will generally be emitted by FBC at levels below the 0.6 lb NO_2/MBtu (258 ng/J) requirement considered in this evaluation. No direct control techniques for oxides of nitrogen have been clearly demonstrated on fluidized-bed combustors to date, although several options are under study.
- The greatest FBC power plant uncertainties presently involve reliability questions - e.g., solids feeding, particulate control (especially for PFBC), material erosion/corrosion/deposition, and process control. The impact of emission standard averaging time basis and system reliability have not been evaluated.
- AFBC and PFBC development programs should focus on more stringent emission standards and their relation to combustor design and operating conditions.
- Advanced FBC sulfur removal concepts, for example, sorbent precalcination, sorbent regeneration, sorbent fines reconstitution, additives for improved sorbent utilization, alternative metal oxide sorbents should be evaluated with respect to more stringent emission standards.

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APPENDIX A

SULFUR OXIDE REMOVAL DATA BASE AND MODEL

SUMMARY

The projections of calcium-to-sulfur molar feed ratio required to achieve any selected degree of desulfurization in a fluidized-bed combustor are derived using a simplified model for fluidized-bed desulfurization, with kinetics rate constants developed using laboratory thermogravimetric (TG) data. For confirmation, where possible, the TG-supported model has been compared with available data from fluidized-bed combustors. The TG data base, the model, and its agreement with fluidized-bed data are discussed. Operating and design conditions under which a typical limestone can be used to maintain 90 percent sulfur removal in AFBC are outlined. Areas for improvement of the projection technique are identified.

INTRODUCTION

The feed ratio of calcium to sulfur required for desulfurization in fluidized-bed combustion depends on the system design, the operating conditions, the particular calcium-based sorbent, and the sorbent particle size used. In assessments of the overall FBC system it has been convenient in the past to select calcium/sulfur feed ratios of about 3/1 for AFBC and about 2/1 for PFBC as base cases, and, over the years, these estimates have acquired a more permanent status, without reference to any particular sorbent, system design, or operating conditions. Desulfurization in the bed, however, is essentially a competition between two processes: once sulfur dioxide is released in the bed it may either remain in the gas and escape or react with calcium oxide (CaO) or carbonate ($CaCO_3$) in the bed to form calcium sulfate ($CaSO_4$). The balance between these processes can be drastically altered by enhancing the probability of escape (shallow bed, high gas velocity) or by changing the reaction rate of the sorbent (e.g., decreasing the sorbent particle size, raising the bed temperature at atmospheric pressure).

In order to predict the calcium-to-sulfur molar rate required for a given level of desulfurization, a model of the desulfurization system that encompasses the effects of the relevant variables is required. Such a model must succeed in correctly projecting the effects of key variables on the system. In particular, since the following effects have been observed in experimental combustors, it should demonstrate

- That an optimum temperature for desulfurization exists at atmospheric pressure using sorbent particle of size 500 μm or larger
- That sorbent utilization improves at higher pressures (∿1000 kPa)
- That there is no marked temperature effect at higher pressures (∿1000 kPa).

Attempts to model fluidized-bed desulfurization to the extent of predicting calcium-to-sulfur molar feed ratios may be: a) statistical correlations based on analysis of fluidized-bed combustion data (Babcock & Wilcox [B&W], Battelle[A2]); b) models which attempt to use sulfation rate data obtained in the laboratory (Argonne National Laboratories [ANL],[A3] Westinghouse,[A4] National Coal Board [NCB][A5]; or c) fundamental models in which the gas/solid reaction model is matched with the fluid dynamics of the bed (no successful model known).

The correlation models are generally untrustworthy outside the range of data correlated, and since the objective is to project desulfurization under novel constraints, they are not useful here. Of the other models, the simplest and most direct is that developed at Westinghouse, and it was used here because it has given satisfactory results in the past.

It may be noted that recent fundamental gas-solid modeling by Hartman and Coughlin[A6] has yielded a sufficiently close prediction of effects observed in practice to be considered as a starting point for the development of a fundamental model: development and analysis of a reasonable body of self-consistent rate data has not yet been reported.

THE WESTINGHOUSE MODEL

The model used here has been described previously (Appendix E of Reference A4). The following assumptions are made in deriving the model:

1. Release of sulfur from coal as sulfur dioxide (SO_2) occurs with equal probability at any place in the bed.
2. Sulfur dioxide released at a distance, $\overline{h}$, below the upper bed surface passes in plug-flow through a homogeneous bed of sorbent of height h. If a mean reaction rate, k, can be assigned to the absorption of gas in the bed, the fraction of the sulfur retained is $1 - e^{-kt}$, where t is the gas residence time in the bed of height h. Integration for all points $\overline{h}$ along the bed height yields an expression for the fractional sulfur removal, R, $R = 1 - \frac{1}{kz} (1 - e^{-kz})$, where z is the residence time of gas entering the bed from below.
3. The mean rate constant for gas absorption in the bed, k, is obtained at any fixed mean utilization of the sorbent from thermogravimetric data reduced to unit gas concentration. The rate constant for any given sorbent type and chemical form will depend on the mean sorbent utilization, sorbent particle size, bed temperature, bed pressure, and gas composition in the bed.

The model of sulfur generation in the bed assumes that sulfur is released from coal with equal probability at all places in the bed. While this is a reasonable general assumption, it implies that, for example, 10 percent of the sulfur is released within the top 10 percent of bed height. Assume that all the sulfur generated in the bed below the top 10 percent is captured with 100 percent efficiency, and the overall desulfurization target is 97 percent. Then the top 10 percent of the bed must capture the sulfur liberated within it with at least 70 percent efficiency, despite the small remaining bed height and effective gas residence time.

As the emission standard becomes more stringent, the reaction constant term, kz, required between SO_2 and CaO increases as the following table shows:

Effective Sulfur Dioxide Removal Standard, %	Rate Constant/Rate Constant at 80% (at constant gas residence time)
80	1
90	2
95	4
97.5	8
98.75	16

The rate constant for limestone sulfation required for varied sulfur removal efficiencies is projected in Figure A1 as a function of gas residence time. In practice, the rate constant could be varied by changing, e.g., the sorbent particle size or bed temperature. Alternatively, bed residence time could be adjusted by varying gas velocity or bed depth.

If all of the sulfur is actually liberated in the lower two-thirds of the bed, then a different picture emerges. The top third of the bed behaves only as an absorber into which SO_2 is fed from the bottom. The reaction rates required for desulfurization with this kind of sulfur generation pattern are illustrated in Figure A2.

A generalized sulfur generation model has been published by ANL,[A3] and this could be incorporated into the model discussed here, if desired. The choice of a particular sulfur generation pattern, however, in any case is speculative.

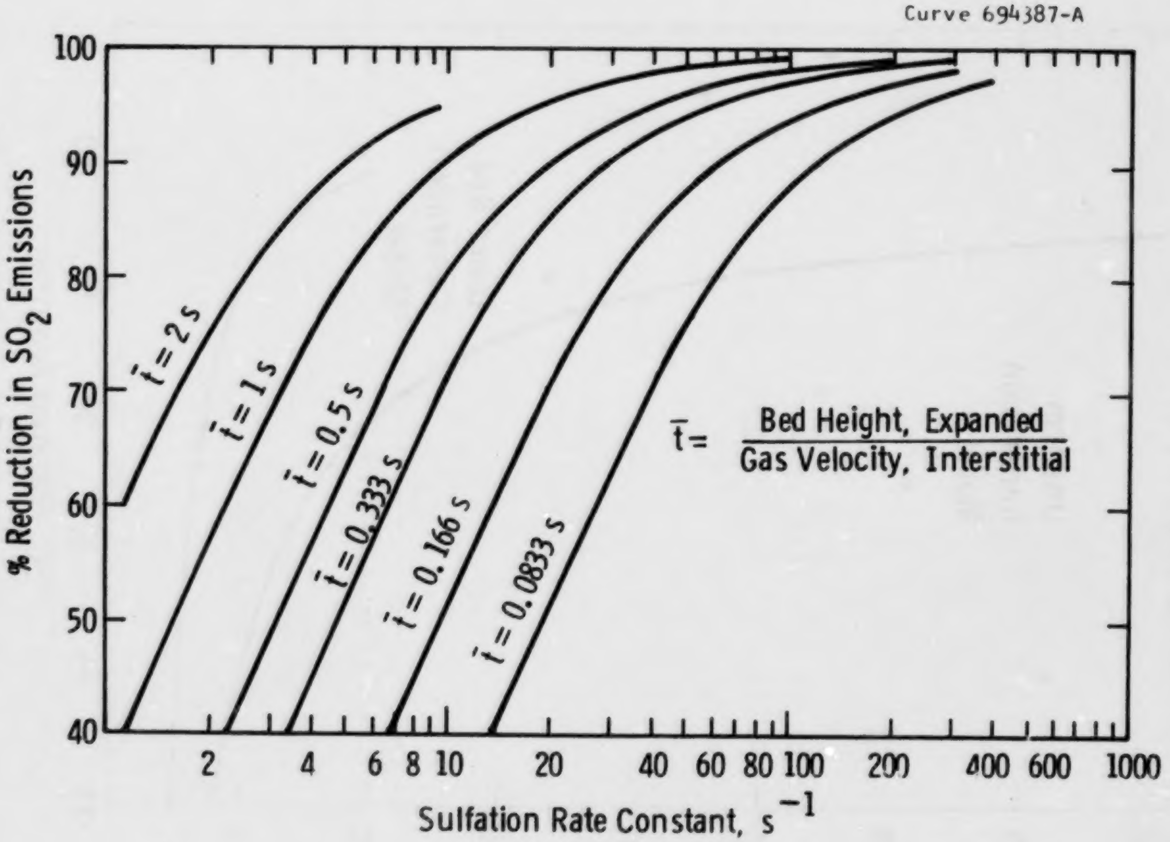

Figure A1 - Required reaction rate as a function of sulfur removal efficiency and gas residence time ($\bar{t}$)

Curve 694388-A

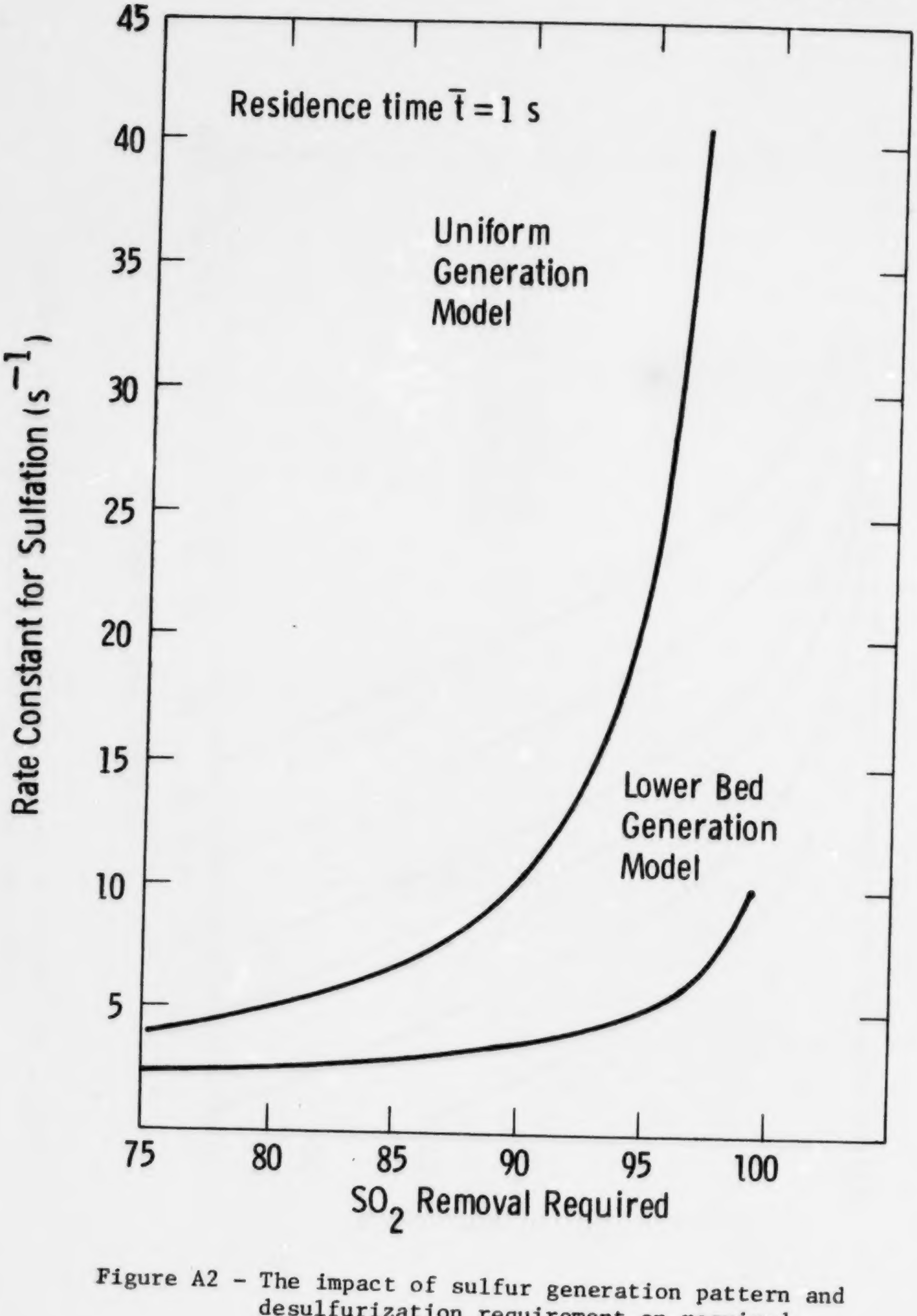

Figure A2 - The impact of sulfur generation pattern and desulfurization requirement on required reaction rate

Operating conditions that would disturb the uniform sulfur generation pattern are: the location of coal feed points and their number, the coal particle size, and the extent of vertical mixing in the operating fluidizing mode. Thus, fine particles of coal fed near the base of a tall bed at low fluidization would probably cause disproportionate release of SO_2 near the base of the bed and improve sulfur removal, relative to the prediction of a uniform sulfur generation model.

TG DATA

The use of TG data to model the fluidized-bed sulfation depends on several assumptions, which are listed below:

1. The rate-limiting process is governed by diffusion within the sorbent itself.
2. The residence time of the sorbent in the bed does not affect its reactivity (although the degree of sulfation does affect reactivity).
3. The fluidized-bed atmosphere surrounding the limestone particles is oxidizing with respect to the $CaS/CaSO_4$ transition.

The implication of the first assumption is that while mass transfer (e.g., of SO_2 from the gas phase to the sorbent particle) in the fluidized bed may be much greater than in the TG apparatus, mass transfer dominates the actual reaction rate over a small extent of reaction. Once the calcium utilization has reached about 10 percent, the rate is usually dominated instead by inter- and intragranular diffusion of the solid. (In same pressurized cases comparatively low mass transfer depresses the observed TG rate over a significant portion of the rate curve). The same internal pore structure must be provided in the sorbent during the TG test as would prevail during a fluid-bed combustor operation. Thus, the calcination conditions under which the sorbent pore structure is generated must be carefully controlled before the TG sulfation experiment if the TG data are to provide a valid simulation of the effect of sorbent utilization on reaction rate.

Concerning the second assumption above, the mean residence time of the sorbent may be much longer in a fluidized bed (up to 24 hours) than in a TG experiment (∿2 hours), unless the latter is carried out under conditions of low-SO_2 concentration (0.05%). The sorbent pore structure may be modified by this long exposure to temperature, and, unfortunately, this may either enhance or retard the sulfation rate, depending on the particular sorbent. This effect is not accounted for in the TG-based kinetic model.

It is often postulated that the existence of local reducing areas in the bed may cause sulfur capture as CaS, which is subsequently oxidized to sulfate. The assumption that this mechanism is unimportant in determining the overall rate of reaction is justified by the fact that oxidation of CaS in limestone is limited in the same way that utilization of CaO is limited in sulfation. Oxidized sulfided limestone contains an inner layer of sulfided stone which has not been oxidized. Since CaS is not found in the product stone from FBC, it is clear that there is no significant additional reaction rate component from sulfidation of the sorbent.

The TG data, which yield rate as a function of sulfur loading, or sorbent utilization, vary greatly according to the sorbent properties and combustor operating conditions. The variables that affect the reaction rate and can be studied on the TGA are

1. Type of sorbent
2. Form of sorbent (e.g., limestone, lime, hydrated lime)
3. Particle size of sorbent
4. Calcination conditions (in particular the temperature and partial pressure of CO_2 during calcination)
5. Residence time (time at temperature)
6. System pressure
7. System temperature
8. Percent excess air in the system
9. SO_2 partial pressure and other gas composition variables.

In making a projection of the effect of operating conditions on the calcium-to-sulfur molar feed ratio, it is advisable to use TG data obtained under experimental conditions that correspond closely to those that will be encountered in the fluidized bed.

For any set of design input:

- Bed height
- Gas velocity
- Bed voidage
- Bed emulsion and bubble phase volume
- Coal:
  - Ash content (heating value)
  - Sulfur content

the rate data can be used in the model to give calcium-to-sulfur molar feed ratio projections. A data base of over 300 atmospheric-pressure and 70 pressurized TG runs exists at Westinghouse over the range of conditions listed in Tables A1 and A2.

BASE OPERATING CONDITION PROJECTIONS

Sorbent feed projections were obtained for the base operating conditions (Table A3) using TG runs 231, 241, P17 and 75-281. The rate data were fitted with polynomial equations for use in the computer generated projections. Figures A3-A6 illustrate the polynomial fits used for TG rate data.

The average rate constant for SO_2 sorption in the bed, K, that is needed to maintain a given level of sulfur retention, R, was calculated at a defined gas residence time ($Z = \frac{\text{expanded bed height}}{\text{interstitial gas velocity}}$) using the equation

$$R = 1 - \frac{1}{KZ}(1 - e^{-KZ}) \quad .$$

Note that K is related to the first-order rate constant for the reaction $CaO + SO_2 + 1/2\ O_2 \rightarrow CaSO_4$, K^1, by

$$K = K^1 \frac{(1 - \gamma)\varepsilon(1 - F)}{\gamma + (1 - \gamma)\varepsilon}$$

where

$K^1 = f(U)$
γ = volume fraction of bed bubbles
ε = bed voidage in emulsion phase
F = fraction of emulsion volume occupied by inerts.

For these projections, the following were assumed:

$$\varepsilon = \varepsilon\ TG = 0.5$$

$$\gamma = 0.5$$

$$F = 0 \quad .$$

In other words, it was assumed that the volume fraction of active emulsion phase in the bed is 0.5 (one-half of the bed volume is tubes, bubbles, and inerts). Thus, the first-order reaction rate constant can be calculated for any sulfur retention level.

The sorbent utilization, U, at which the reaction rate constant, K^1, applies was determined from TG data. A mass balance on the TG system gives the molar rate of reaction, $\frac{dU}{dt}$, in terms of the reaction note constant as

$$\frac{dU}{dt} = \frac{K^1\ C\ \varepsilon_{TG}}{p(1 - \varepsilon_{TG})}$$

where

p = solids density, mole Ca/cc
C = mole SO_2/cc in TG reaction gas.

The polynomial equations representing TG rate data $U = f(\frac{dU}{dt})$ were then used to calculate the sorbent utilization. The required calcium-to-sulfur molar feed ratio, C, is then defined by

$$C = \frac{R}{U} \quad .$$

The dolomite calcium-to-sulfur ratio projections in Table A3 for the PFBC case were derived from Figure A5 and A6, using available TG data that were not at the exact conditions of the base PFBC case. As indicated,

Table A1

RANGE OF ATMOSPHERIC TG SULFATIONS

Stone(s)	Size, U.S. Mesh	% Excess Air	Temp., °C	Comment
Ames, Brownwood, 1359, Greer, Carbon 1337, Western, Mississippi Bellefonte	16/18 5–100/200	20	815	Varied calcination
Greer Ames	16/18 35/40	20 20	780–950 750–940	Varied calcination
Ames, Brownwood Greer, Carbon, 1359	35/40	20	900	Calcined at 900°C in 60% CO_2
Ames, 1359, Greer, Carbon	5/6–35/40	20	850	Scattered particle sizes
1359 (2263)*	16/18	20	815	
Carbon	-325	20	800-950	
Tymochtee	16/18	2-16% O_2	815	
Canaan, Kaiser	100/200	20	815	Calcined at 900°C in 60% CO_2

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*Residence time varied (0.1-0.5% SO_2 in sulfating atmosphere).

Table A2

RANGE OF PRESSURIZED TG SULFATIONS, 10 atm

Stone	Size, U.S. Mesh	% Excess Air	Temp., °C	Comment
Canaan	35/40	300	843	
1337	35/40	300	843-1010	
1337	8/10	20	900	Half-calcined
Greer	35/40	300	843-1010	
Greer	6/8	300	900-1010	
Greer	40/100	300	900-1010	
Greer	100/200	300	950-1000	
Greer	16/18	15-100	815	Carbonated
Greer*	14/18	20	815	FB Calcine 815, 15% CO_2
Tymochtee	16/18	Calc. 815 (0.15 atm CO_2) 0.75-16% O_2	815	
1359 (2263)*	16/18	20	815	

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*Residence time varied (0.05-0.5% SO_2 in sulfating atm).

Table A3

FBC PLANT OPERATING CONDITIONS

	AFBC		PFBC	
Pressure, kPa	101		1013	
Bed Temperature, °C	840		950	
Excess Air, %	20		20	
Sorbent Type	Limestone		Dolomite	
Sorbent Diameter, μm	1000	500	2000	500
Bed Velocity, m/s	3.05	1.83	1.52	1.52
Bed Depth, m	1.22	1.22	3.05	3.05
Ca/S for 1.2 lb/MBtu SO_2 Emission[a]	5.5	2.4	2.0	1.2
Ca/S for 90% Sulfur Removal Efficiency	7.0	2.9	4.5	1.7

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[a] 83% sulfur removal for 4 wt % sulfur coal.

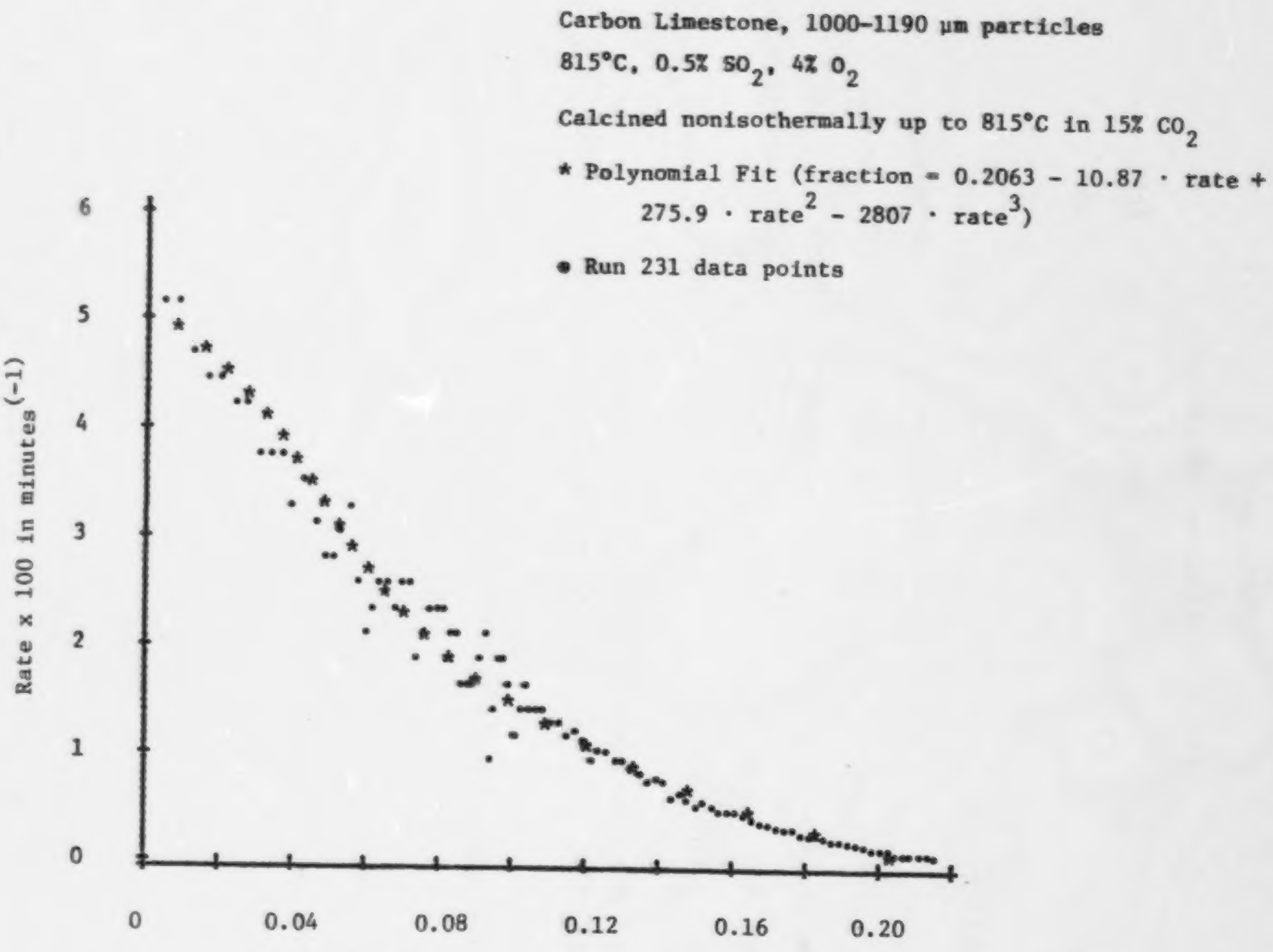

Figure A3 - Polynomial fit of TG rate data from run 231

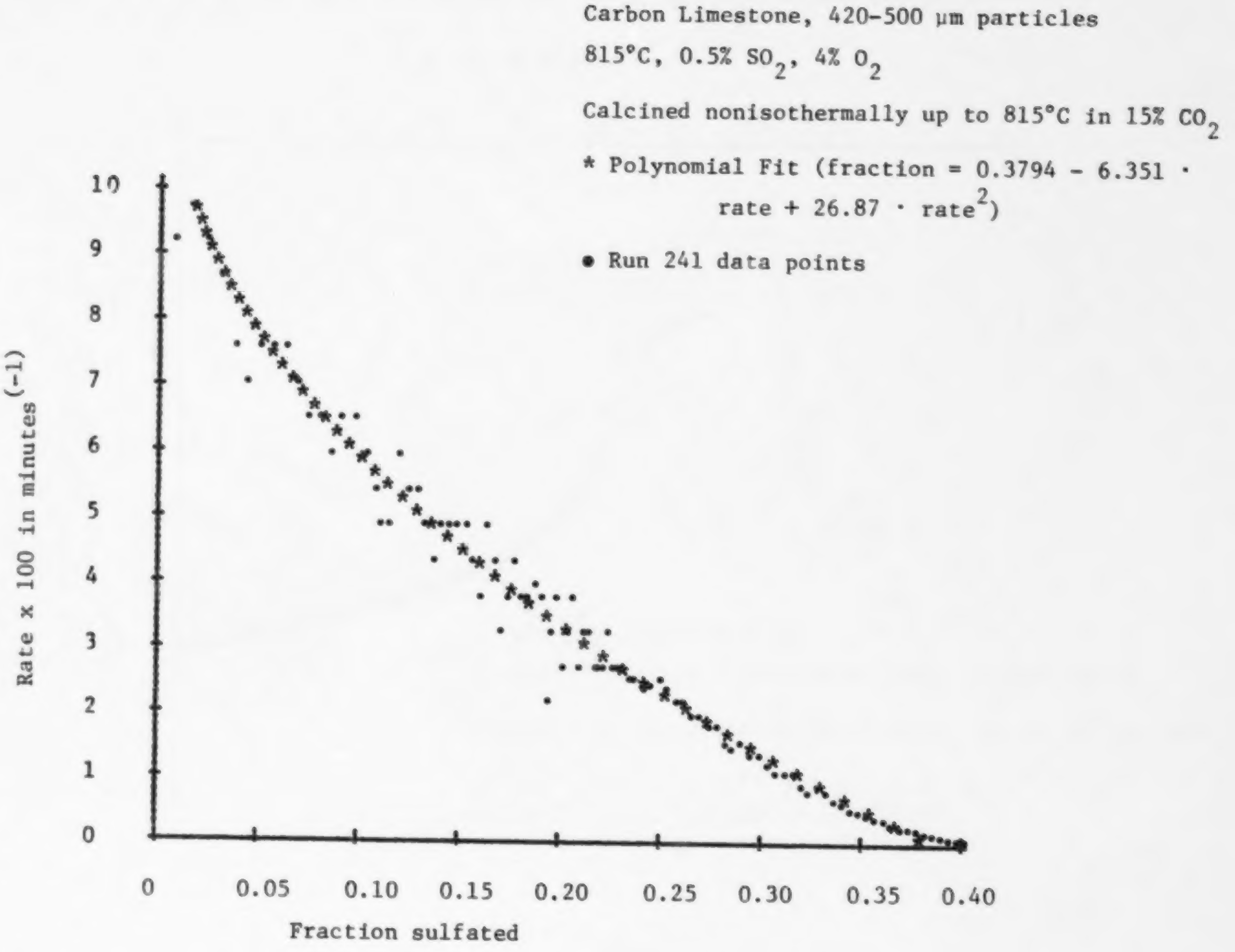

Figure A4 - Polynomial fit of TG rate data from run 241

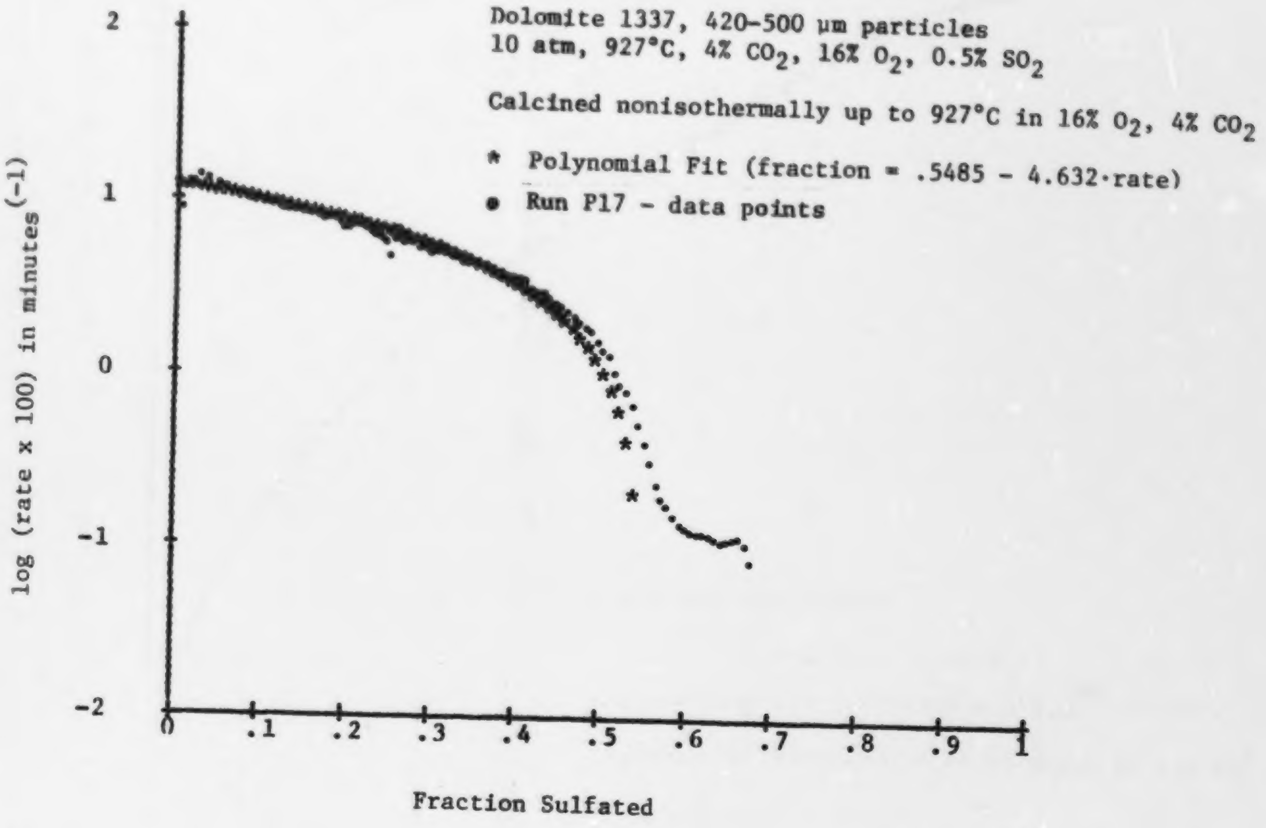

Figure A5 - Polynomial fit of TG rate data from Run P17

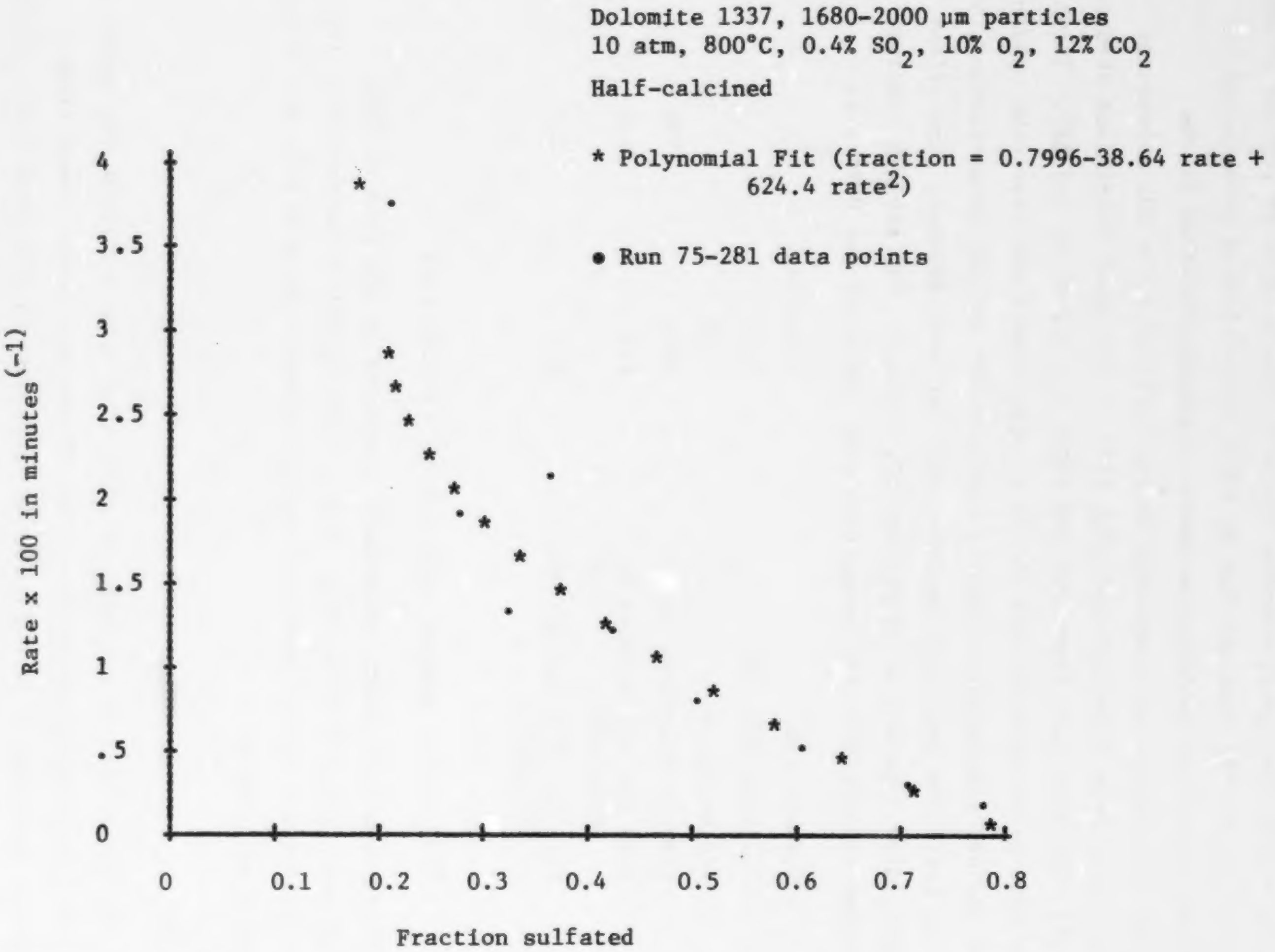

Figure A6 - Polynomial fit of TG rate data from run 75-281

the run P17 TG data in Figure A5 are for 300 percent excess air (16 percent O_2 in the flue gas), whereas the base case assumes 20 percent excess air; the run 75-281 data are for an 800°C half-calcined dolomite of 2000 µm. Since the technical/economic analysis presented in the body of this report was completed in the basis of Table A3, subsequent TG runs have been made with dolomite 1337 at the exact conditions assumed for the PFBC base case (runs P96 and P104, see Figure A7 and A8). It was found that using runs P17 and 75-281 to approximate the base case conditions was not completely accurate; both the 500 µm and, in particular, the 2000 µm particles were more reactive than had been assumed. The effect is especially apparent at 90 percent SO_2 removal. The revised calcium-to-sulfur projections (for comparison with Table A3) for PFBC are:

Sorbent type	Dolomite	
Bed volocity, m/s	1.52	
Bed depth, m	3.05	
Sorbent diameter, µm	2000	500
Ca/S for 1.2 lb/MBtu SO_2 emission[a]	1.9	1.1
Ca/S for 90% sulfur removal efficiency	2.8	1.3

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[a] 83% sulfur removal with a 4 wt % sulfur coal

If the technical/economic assessment discussed in the body of this report were repeated using these new calcium-to-sulfur projections, the PFBC cases would appear more economically attractive than they did in the original assessment.

CONFIRMATION OF THE MODEL

The model used to project the effects of more stringent SO_2 standards has been used previously to show that TG data accurately demonstrate important features of desulfurization phenomena in fluidized beds. Desulfurization phenomena that have been observed in fluidized beds and demonstrated on the TG include:[A4]

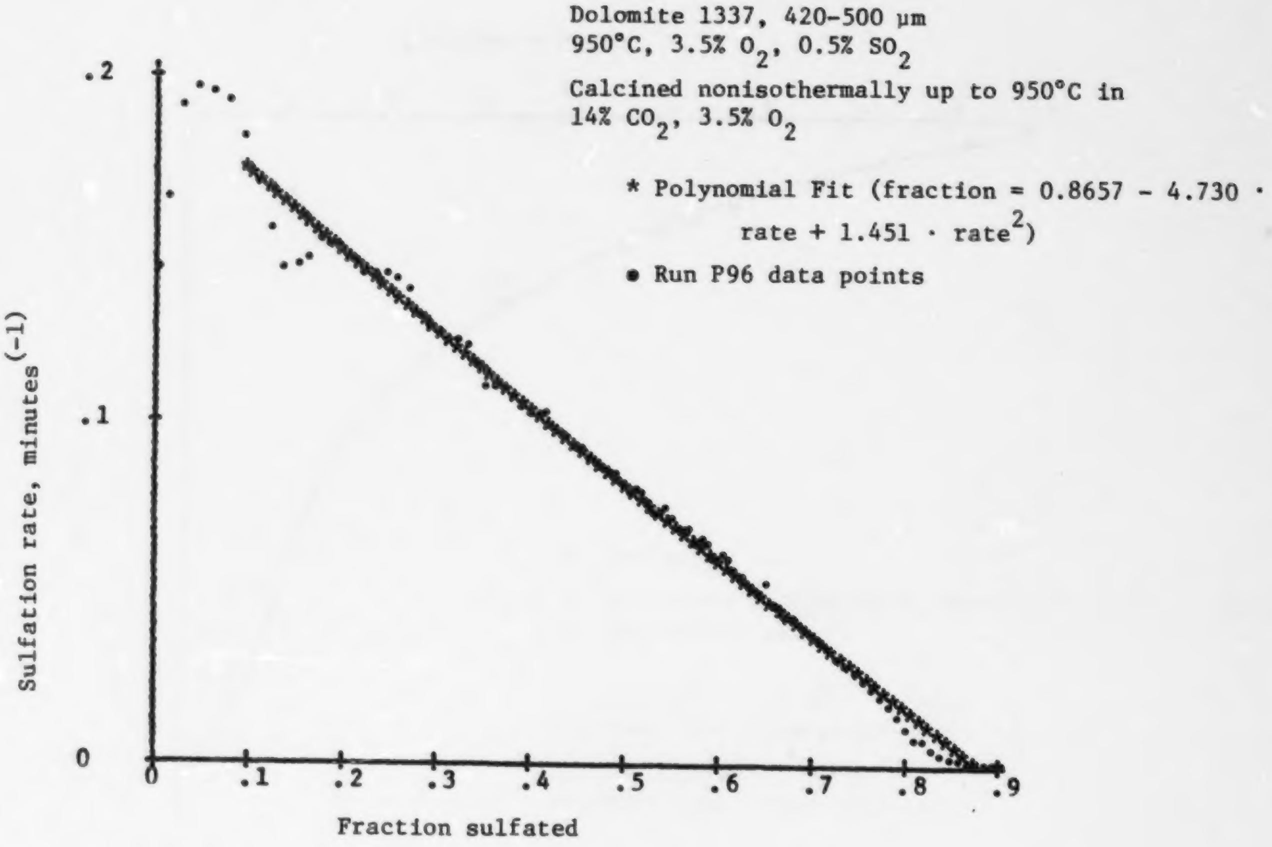

Figure A7 - Polynomial Fit of TG Rate Data from Run P96

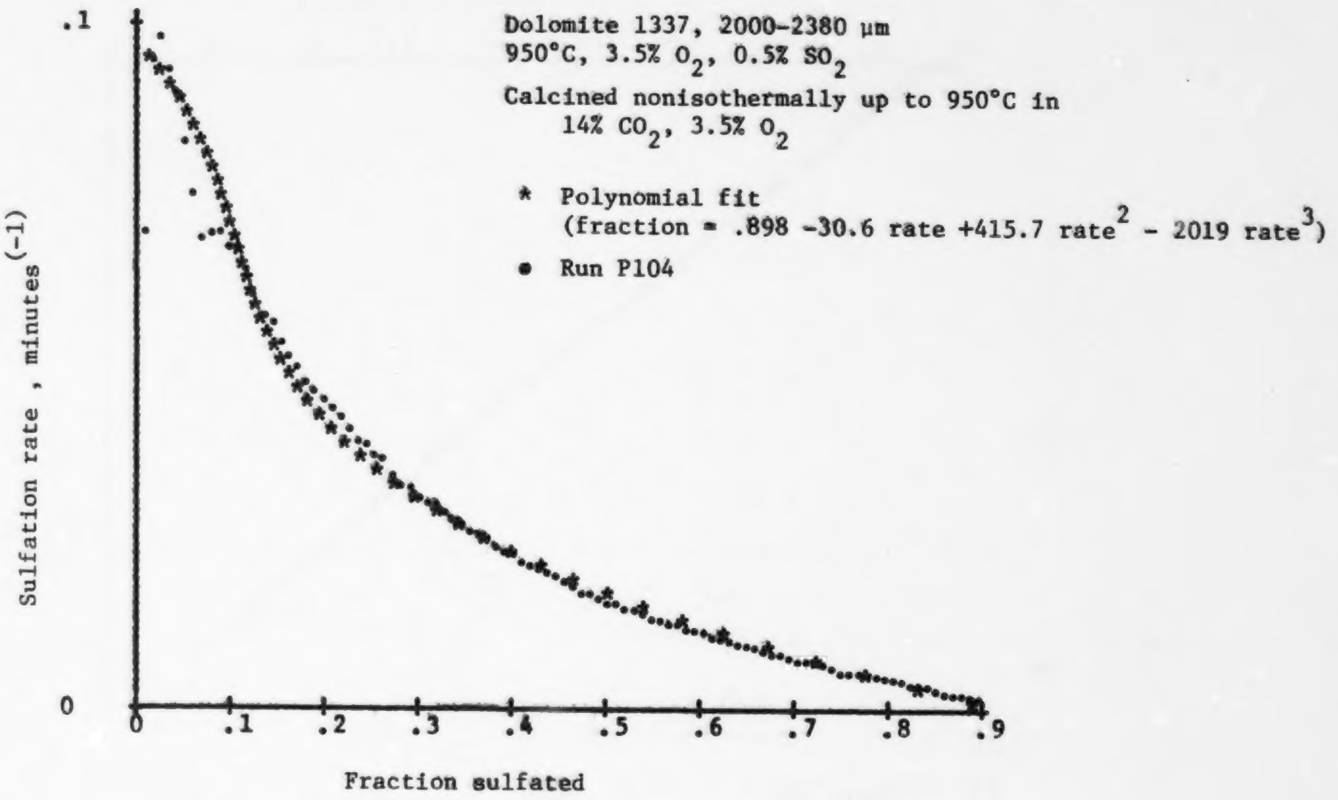

Figure A8 - Polynomial Fit of TG Rate Data from Run P104

- The occurrence of an optimum temperature for desulfurization in AFBC with sorbent particle sizes of greater than 500 μm
- No marked temperature effect at higher pressures (∿1000 kPa)
- Improved sorbent utilization at higher pressures (∿1000 kPa)
- Improved sorbent utilization with precalcination.

The specific data that the TG has been used to model include data from the NCB, ANL, B&W, Pope, Evans and Robbins (PER), and Westinghouse.

Model projections of the calcium-to-sulfur molar feed ratios required for various levels of desulfurization in AFBC, as a function of limestone type, are compared to the data collected from the ANL and British Coal Research fluid-bed units for limestone 1359(A2) in Figure A9. Conditions for the fluid-bed runs were:

1 atm, limestone 1359
490-630 μm particles in the feed
2.6-2.8 ft/s
788-798°C
2 ft bed height
3% O_2, 15% CO_2 in the flue gas.

To obtain the projections, TG rate data from sulfation at 815°C in 0.5% SO_2, 4% O_2, and N_2 were utilized. The sulfations were carried out with 420 to 500 μm particles of limestone, calcined at 815°C in 15% CO_2 and nitrogen. The gas residence time (as determined by input bed height and velocity) was 0.66 s. ANL operated with a gas residence time of 0.74 s. This longer residence time may account for the slightly lower calcium-to-sulfur molar ratio requirements in the ANL 1359 data.

Projections of the calcium-to-sulfur molar feed ratios required for various levels of desulfurization in PFBC are compared with data collected from the Exxon miniplant(A8) in Figure A10. It should be noted that the scatter of miniplant data is partially due to the range of operating conditions chosen for the plant.

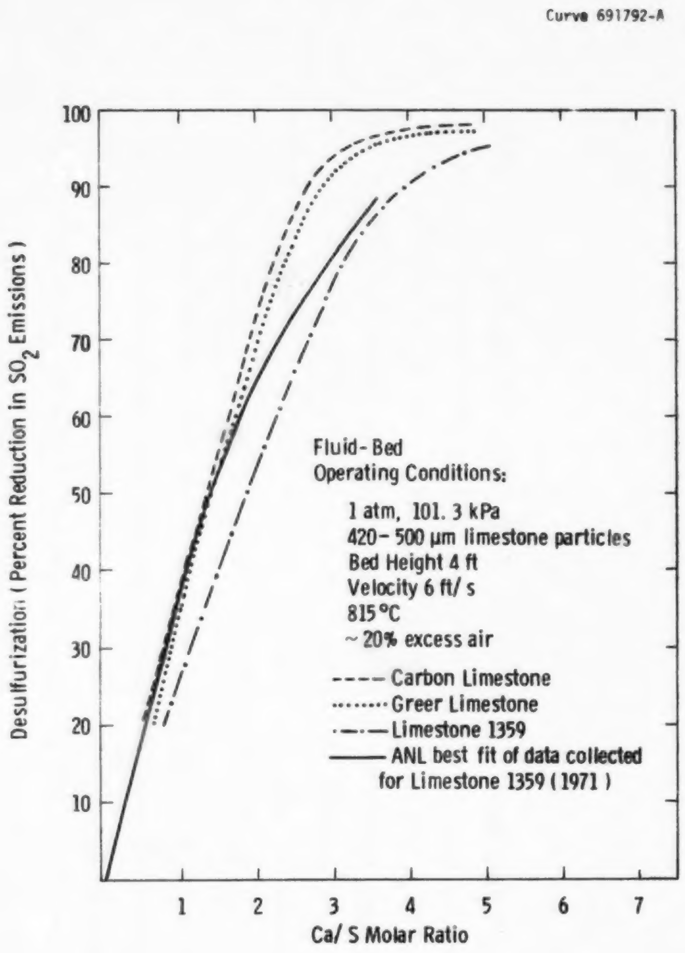

Figure A9 - Predictions of the Ca/S feed ratios required for desulfurization using Westinghouse TG data for three sorbents

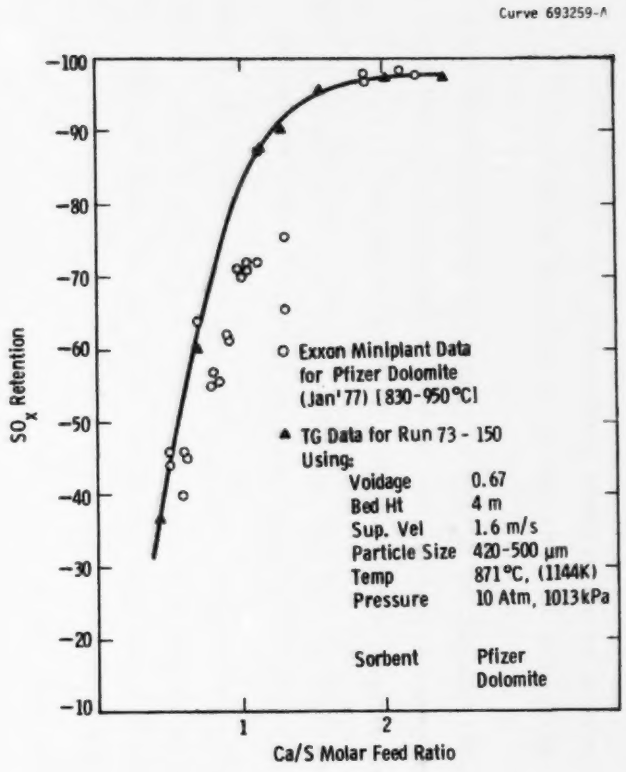

Figure A10 - Sulfur removal projected from TG data and measured on the Exxon miniplant

VALIDITY OF THE ASSESSMENT PROJECTIONS

A computerized file of fluid-bed combustor data from PER, the NCB, ANL, Exxon, and B&W was searched for combustor runs carried out at operating conditions near the base case conditions for the technical/ economic assessment. The limited bench-scale and pilot plant data available at the base case conditions indicate that the TG projections are representative of the fluid-bed data.

Atmospheric combustor data at operating conditions near the base case are given in Tables A4 and A5. Note that, under base case conditions, 80 percent sulfur removal has not been obtained at a calcium-to-sulfur ratio of less than four with 1000 μm particles. Figure A11 compares the calcium-to-sulfur feed requirements for desulfurization from Table A4 with the TG projections for 1000 μm particles. The 500 μm data combustor runs are plotted in Figure A12 with TG projections. The scatter of combustor data is partially due to the range of operating conditions plotted, i.e., 799 to 899°C temperatures, and varied sorbent types.

Pressurized combustor data at operating conditions near the base case with a defined sorbent particle size were not in the data file. Recent results from the Exxon miniplant,[A9] however, indicate that greater than 97 percent sulfur removal can be obtained with calcium-to-sulfur molar feed ratios of less than 2. This is in line with PFBC projections for the base case.

DESIGN CONDITIONS FOR 90 PERCENT SULFUR REMOVAL IN AFBC

The Westinghouse model was used to project the design and operating conditions required to maintain 90 percent sulfur removal in AFBC. TG rate data from finely divided particle size fractions of limestone, with diameter, d, were used to generate calcium-to-sulfur molar ratio requirements for 90 percent sulfur removal as a function of gas residence time

Table A4

FLUID-BED COMBUSTOR DATA AT CONDITIONS NEAR THE 1000 μm AFBC BASE CASE PROJECTIONS

Study	Sorbent	Particle Size, μm	Bed Temperature, °C	Approximate Gas Residence Time, s[a]	Ca/S Molar Ratio	% SO_2 Removal
Continuous-Feed Unit Tests:						
ANL(A10)	1360	1000-1400	871	0.42	2.5	72
ANL(A10)	1360	1000-1400	871	0.42	4.2	86
ANL(A10)	1359	1400	871	0.67	1.6	18
NCB(A11)	18	<3175	777	0.38	2.9	73
B&W(A12)	Carbon	<6350	819	0.19	2.8	58
Batch Unit Tests:						
PER(A13)	1359	840-2381	860	0.14	16.0	93
Exxon(A14)	1359	930	871	0.25	6.8	90
Exxon(A14)	1359	930	871	0.25	5.4	90
Exxon(A14)	1359	930	982	0.25	5.1	90
Exxon(A14)	1359	930	927	0.28	5.9	90
Westinghouse Projection:						
	Carbon	1000-1190	840	0.4	5.5	83
	Carbon	1000-1190	840	0.4	7.0	90

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[a] Gas Residence Time = $\frac{\text{Bed height expanded}}{\text{Superficial Velocity}}$.

Table A5

FLUID-BED COMBUSTOR DATA AT CONDITIONS NEAR THE 500 μm AFBC BASE CASE PROJECTIONS

(Continuous-Feed Unit Tests)

Study	Sorbent	Particle Size, μm	Bed Temperature, °C	Approximate Gas Residence Time, s[a]	Ca/S Molar Ratio	% SO_2 Removal
ANL(A3)	1360	630	871	0.7	2.3	74
ANL(A3)	1359	490	815	0.7	2.5	91
ANL(A3)	1359	490	843	0.7	2.5	91
ANL(A3)	1359	490	871	0.7	2.5	86
ANL(A3)	1359	615	899	0.7	2.6	79
ANL(A3)	1359	615	899	0.7	2.3	83
ANL(A3)	1359	615	899	0.7	2.5	79
ANL(A15)	1359	609	843	0.63	11.9	90
ANL(A15)	1359	609	843	0.66	5.4	88
NCB(A11)	UK	537	799	0.7	3.2	94
NCB(A11)	UK	537	799	0.7	1.8	81
NCB(A11)	UK	537	799	0.7	2.9	97
NCB(A11)	UK	537	799	1.0	1.8	85
NCB(A11)	18	453	848	0.7	5.7	97
NCB(A11)	18	453	848	0.7	6.0	100
NCB(A11)	UK	537	799	1.0	2.6	84
NCB(A11)	UK	537	799	1.0	3.1	88
Westinghouse Projections						
	Carbon	420-500	840	0.7	2.4	83
	Carbon	420-500	840	0.7	2.9	90

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[a] Gas Residence Time = $\frac{\text{Bed height expanded}}{\text{Superficial velocity}}$.

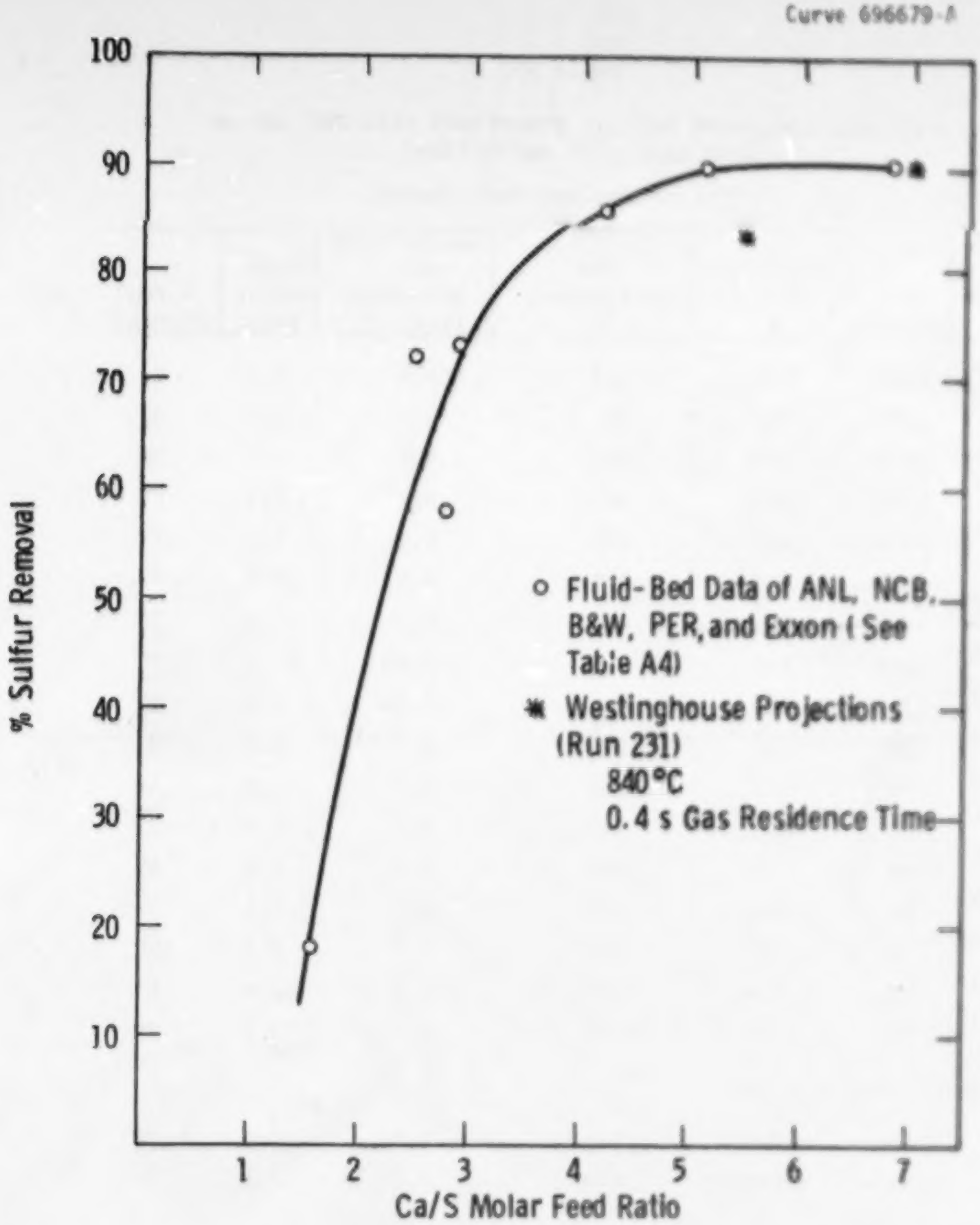

Figure A11 - Fluid-bed combustor data at conditions near the 1000 μm AFBC base-case projection conditions

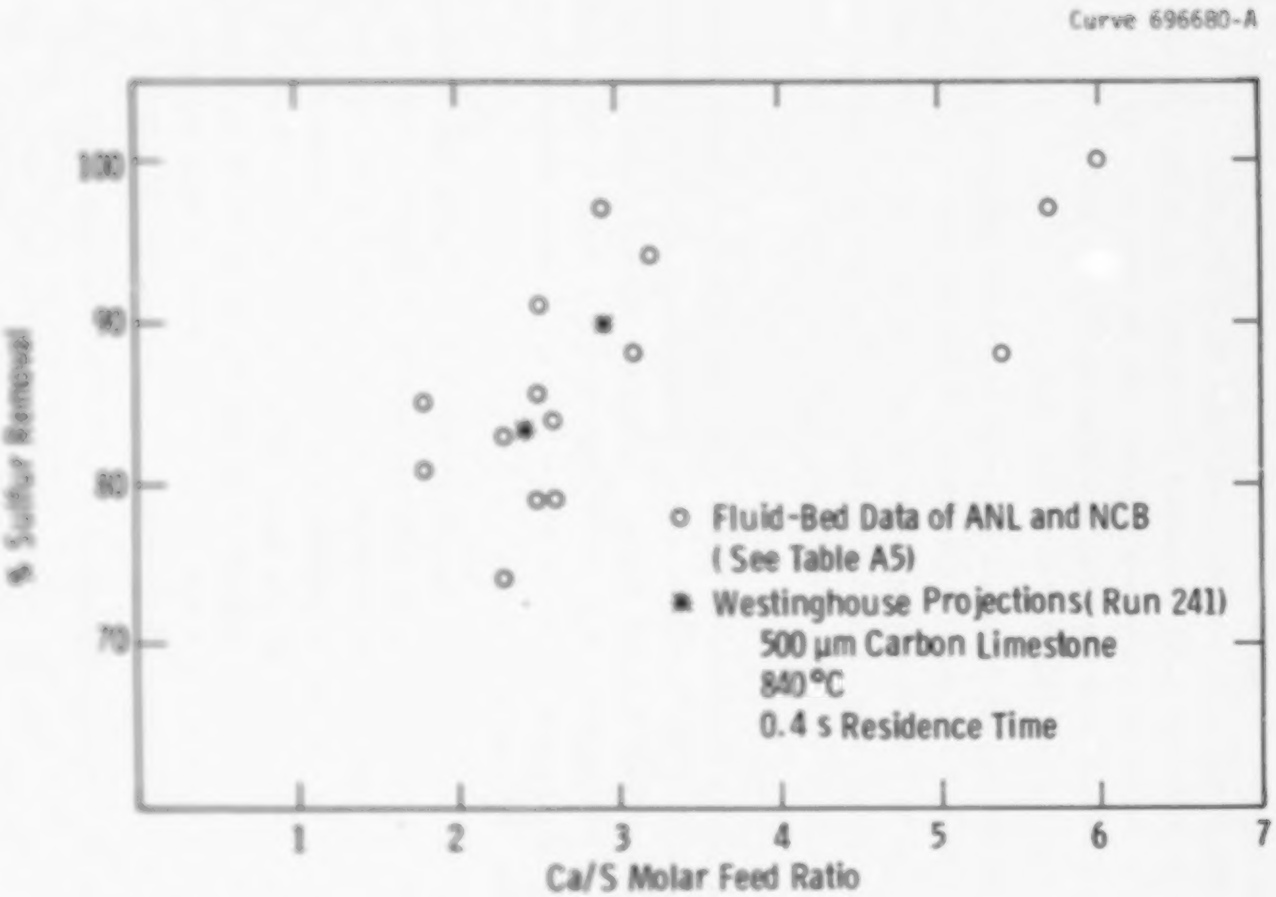

Figure A12 - Fluid-bed combustor data at conditions near the 500 µm AFBC base-case projection conditions

and sorbent particle size.* The projections are valid for a once-through AFBC system using no sorbent enhancement options (such as precalcination), operating at 815°C with 20 percent excess air. For these calculations, it was assumed that no sulfur is released from the carbon burnup cell and that fines' residence time does not limit their utilization. The technique by which the model was utilized to make these projections was described in detail in a previous section. The TG data used, and the polynomials derived from the data, are shown in Figures A13 through A18 (TG data from Figures A3 and A4 were also used).

Results for Carbon limestone are shown in Figure A19. The figure shows that for a 0.5 s gas residence time, 90 percent sulfur removal can be maintained with a calcium-to-sulfur molar feed ratio of 5/1 when the limestone particle size is less than 1000 µm. The calcium-to-sulfur molar ratio can be reduced to 3/1 if the particle size is less than 600 µm.

Using 500 µm particles of Carbon limestone, a 5/1 calcium-to-sulfur molar feed will be sufficient to remove 90 percent of the sulfur generated with a gas residence time of <0.1 s. Increasing the gas residence time to 0.18 s will reduce the required calcium-to-sulfur ratio to 3/1. Further increasing the gas residence time to 1 s will only reduce the calcium-to-sulfur molar feed ratio to 2.5/1.

It should be noted that the proper selection of design and operating conditions is an important, but limited, means of reducing calcium consumption. For example, 500 µm Carbon limestone particles will not meet 90 percent sulfur removal requirements with a calcium-to-sulfur molar feed ratio of 2/1 at any operating or design condition. The TG rate curve in Figure A20 shows that at 30 percent sulfation of 500 µm Carbon limestone, the rate drops rapidly as diffusion of SO_2 through product sulfate limits the stone's utilization. If greater than 60 percent

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*It was assumed that projections developed from a particle size, d, are representative of a feed particle size distribution with a mean diameter, $\bar{d}$. Past projections based on this assumption have been accurate. However, distributions in which d does not represent $\bar{d}$ may exist. No systematic study on representing feed particle size distributions has yet been done.

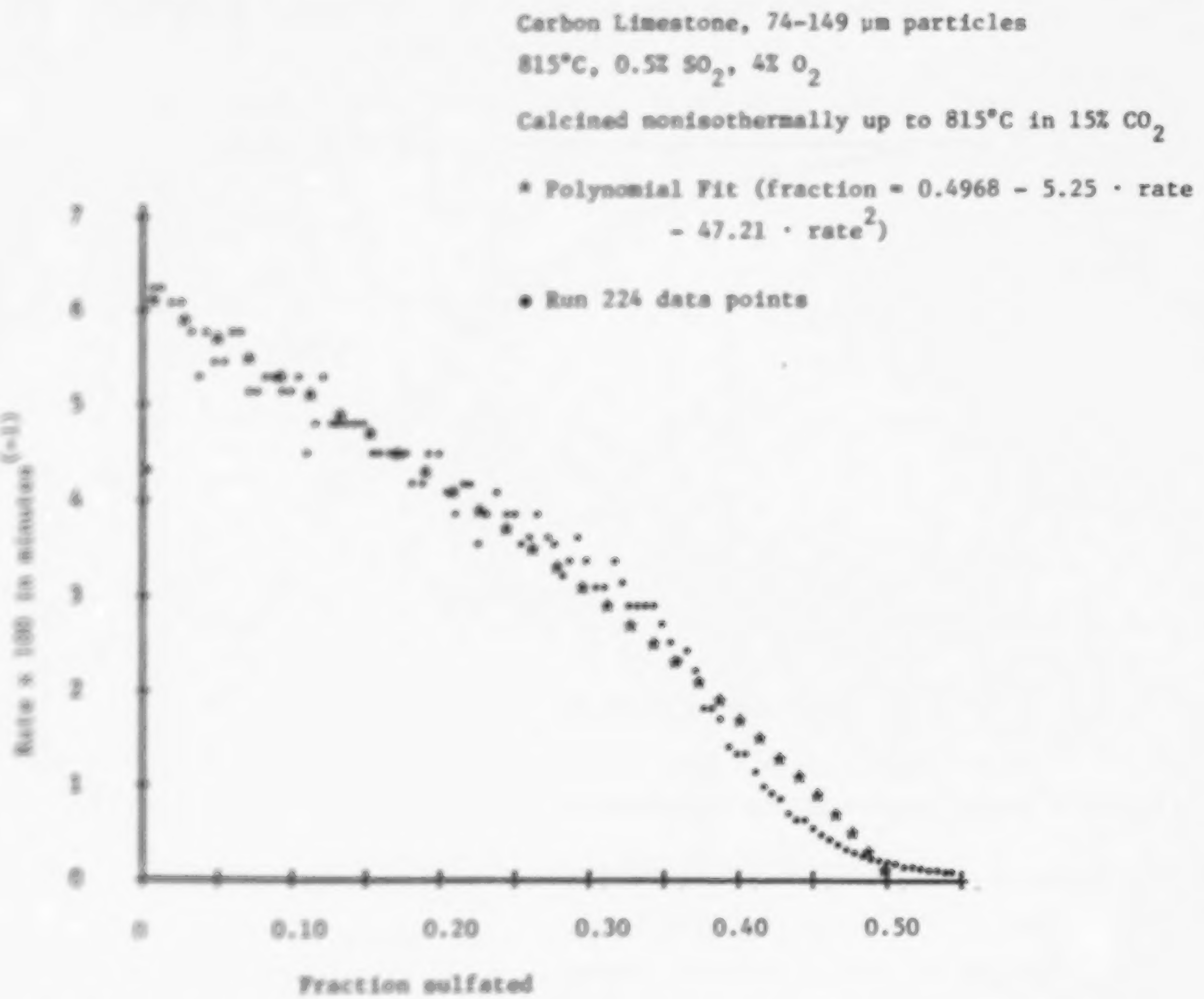

Figure A13 – Polynomial fit of the TG rate data from run 224

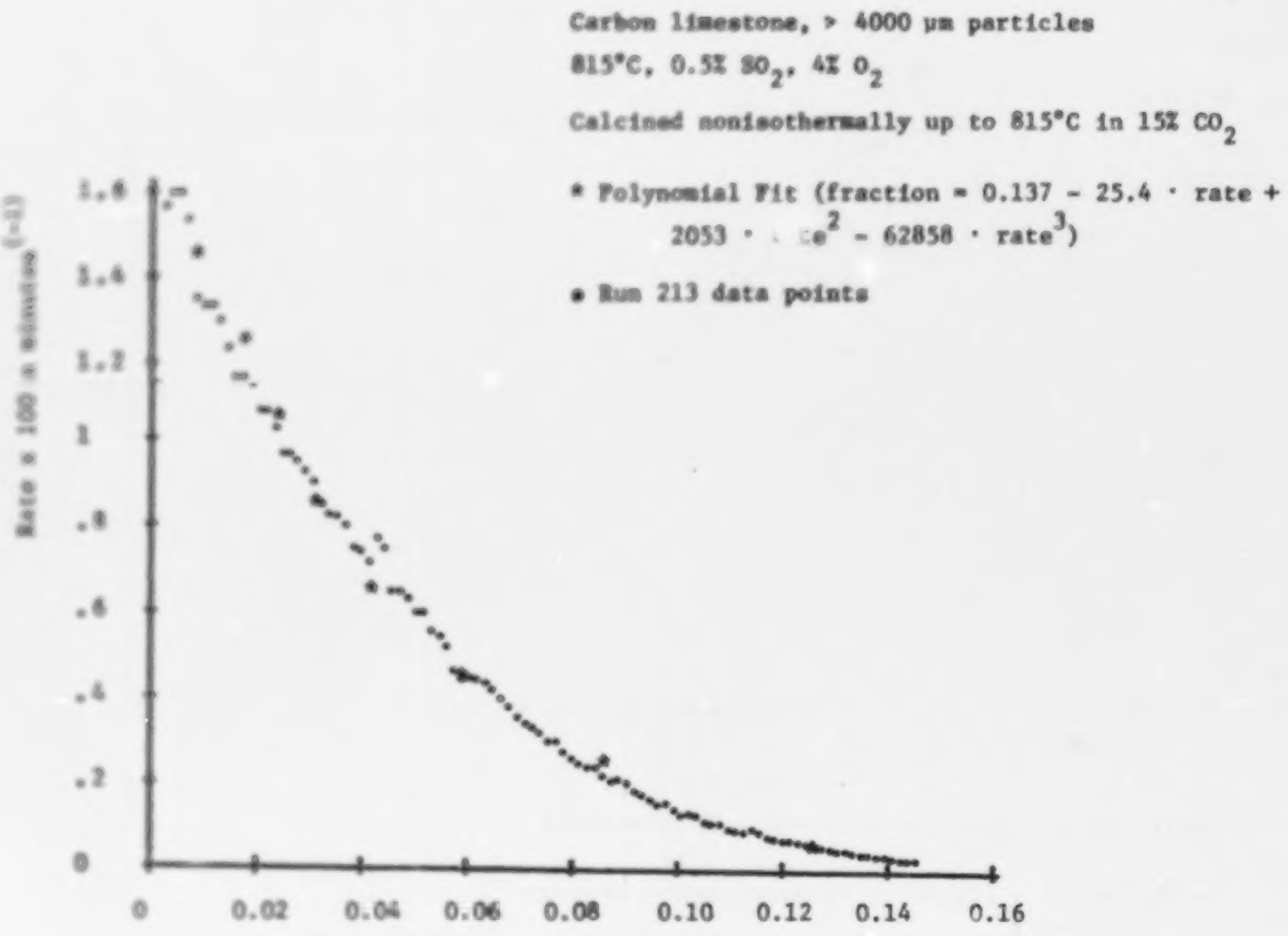

Figure A14 - Polynomial fit of the TG rate data from run 213

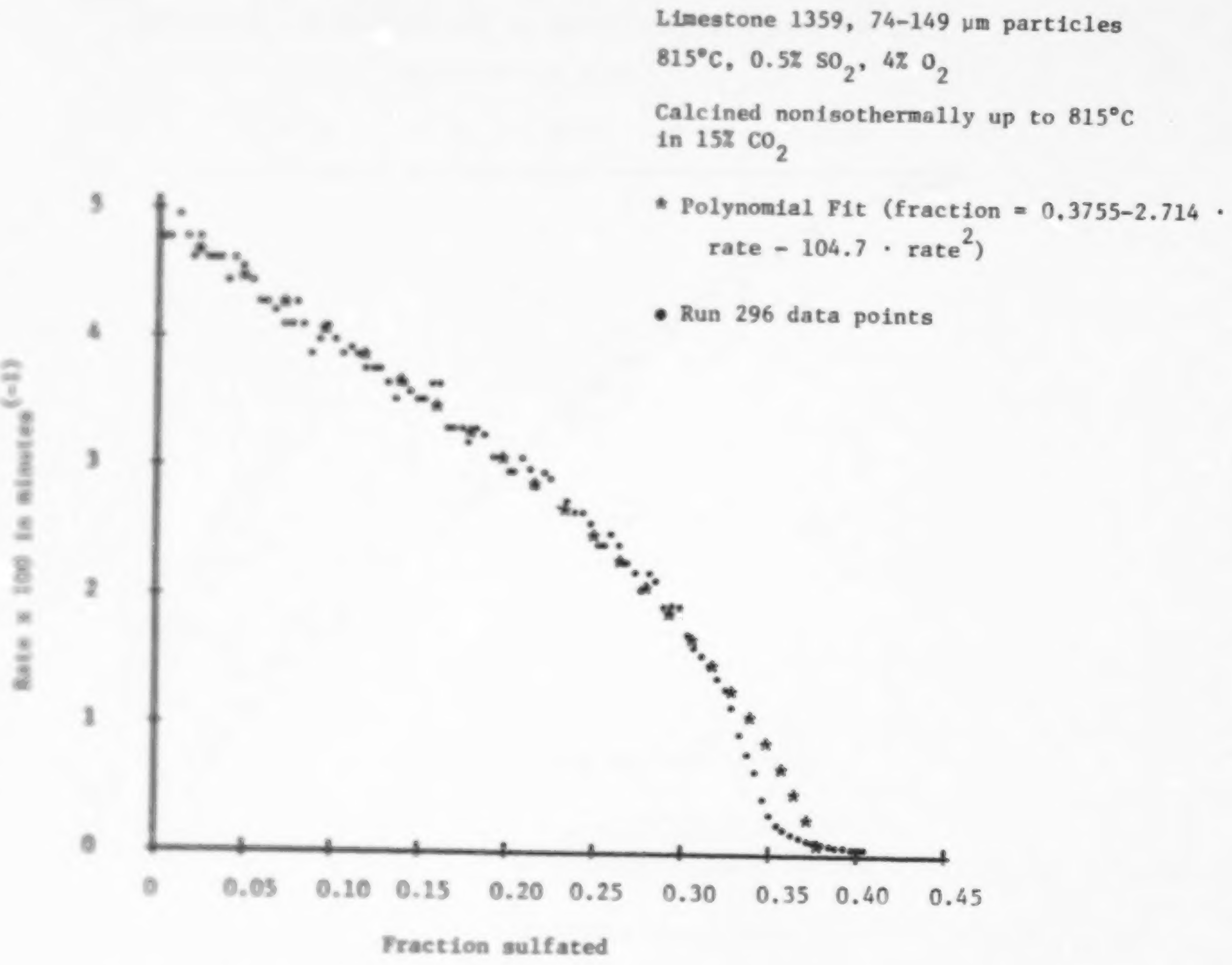

Figure A15 - Polynomial fit of the TG rate data from run 296

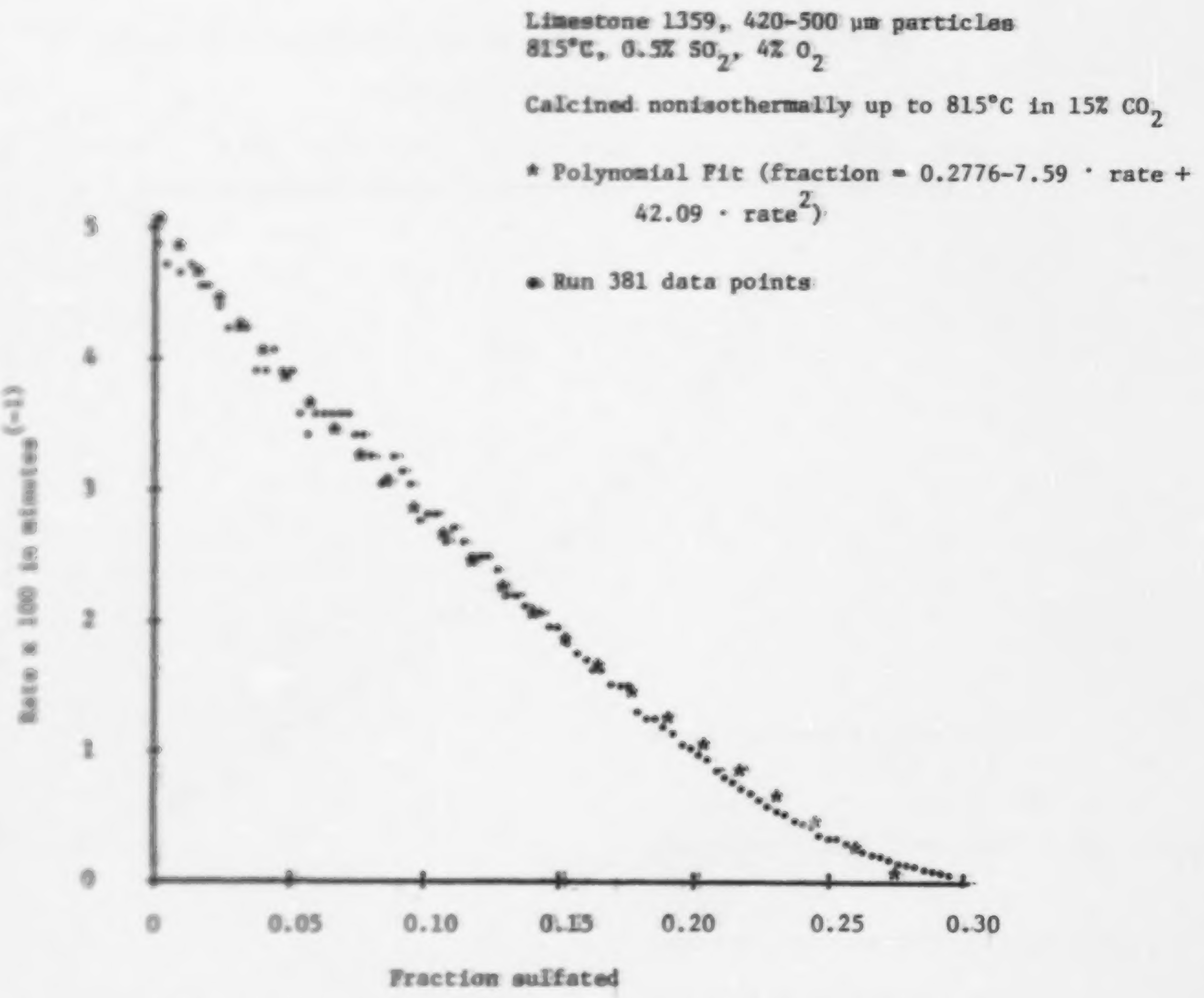

Figure A16 - Polynomial fit of the TG rate data from run 381

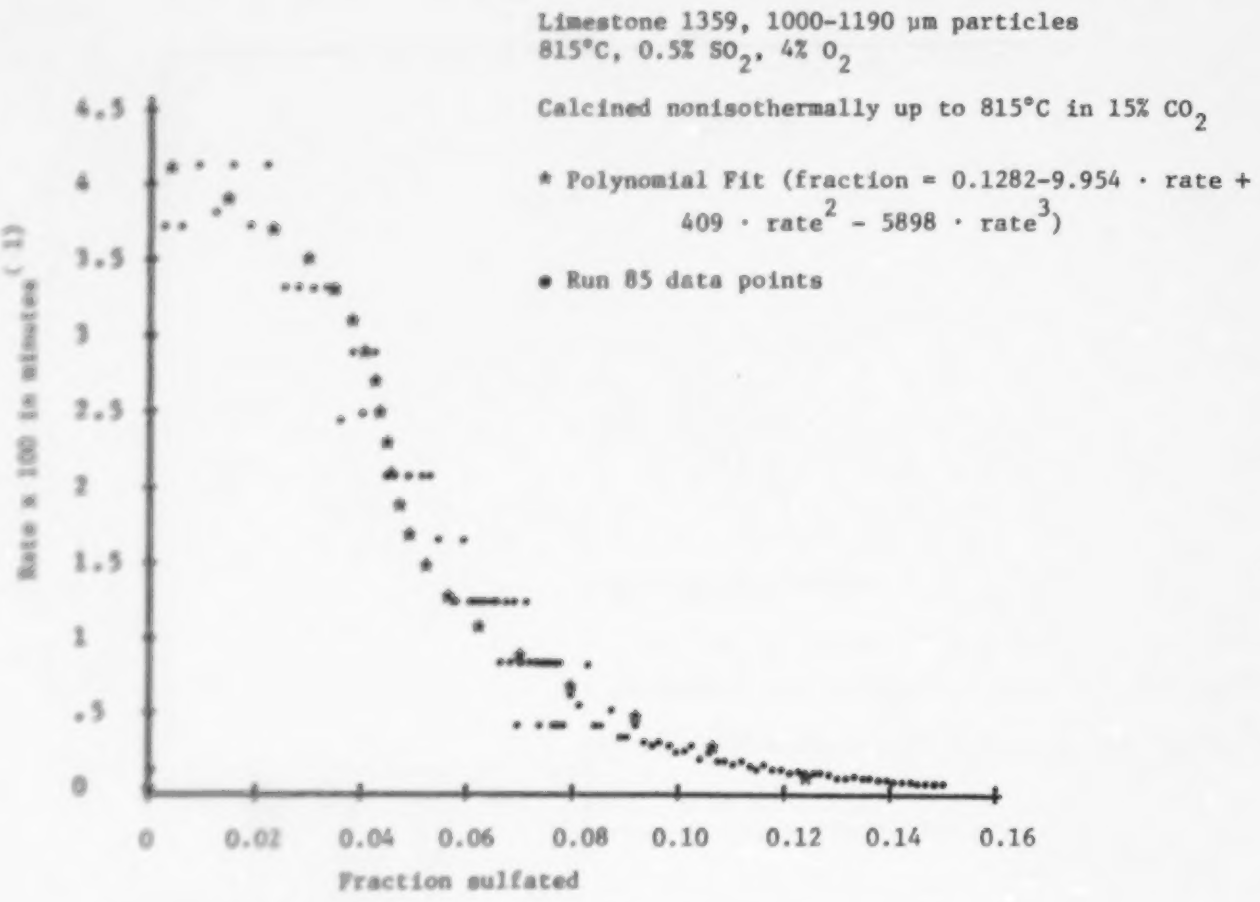

Figure A17 - Polynomial fit of the TG rate data from run 85

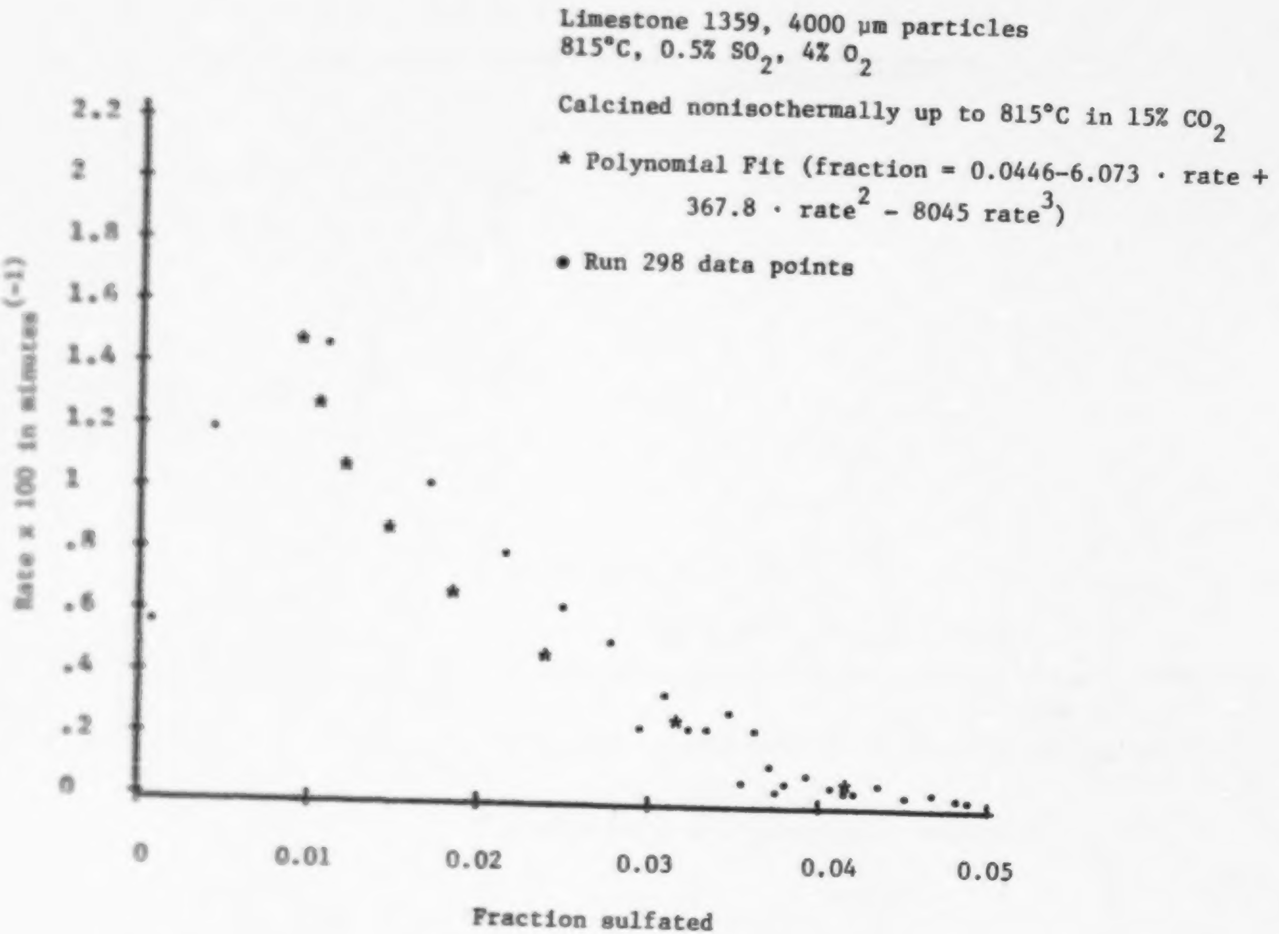

Figure A18 - Polynomial fit of the TG rate data from run 298

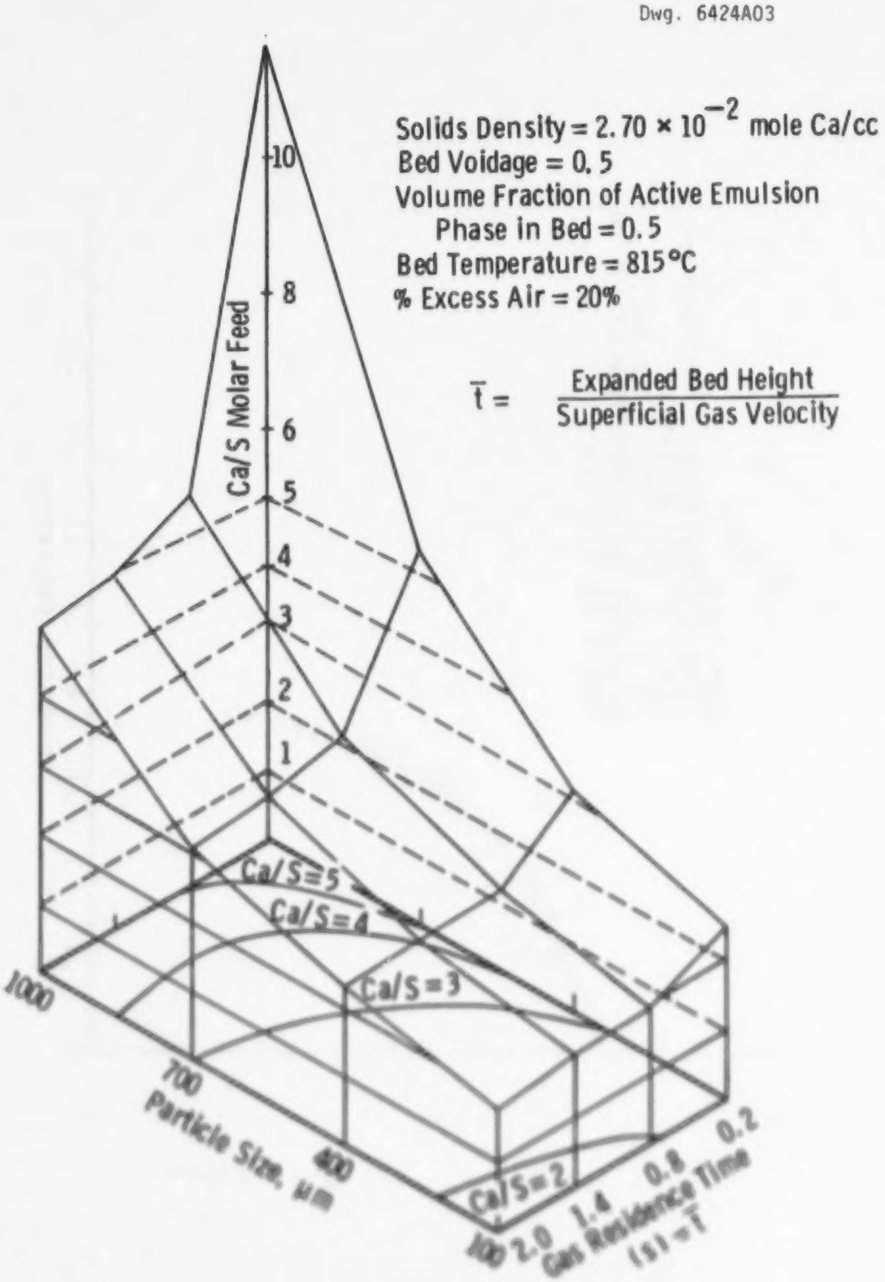

Figure A19 - Ca/S molar feed required to maintain 90% sulfur removal in AFBC with Carbon limestone

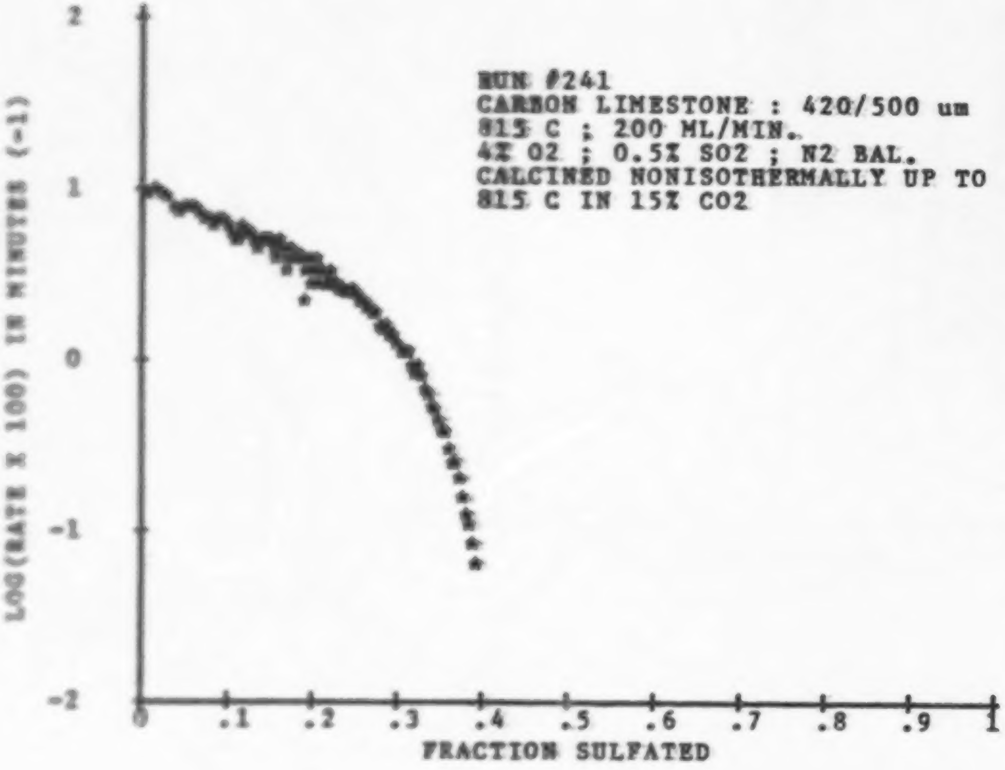

Figure A20 - The sulfation rate of Carbon limestone

utilization of a 500 μm sorbent is required, a calcine with a more open pore structure must be used. The pore structure of Carbon limestone can be modified by a sorbent activation technique, or another sorbent with a greater utilization potential can be chosen.

The variation in calcium feed requirements with sorbent type is illustrated by comparing Figures A19 with the results for Grove limestone (limestone 1359) in Figure A21. Grove limestone may require twice the calcium molar feed as Carbon limestone to meet 90 percent sulfur removal requirements.

AMENDMENTS TO THE MODEL

Future amendments to be made on the model include a) integration of sorbent particle size distribution, b) including the impact of attrition and elutriation on particle size distribution, and c) discerning the effect of different mass transfer rates on the TG and in fluid beds.

Instead of using a TG cure based on one particle size representing the estimated average particlesize in the bed, it would be preferable to use a composite TG curve representing the fuel particle size distribution within the bed. Westinghouse has already calculated such a composite curve for one particular case - the atmospheric-pressure experiments of B&W,[A12] with exellent results.

The Westinghouse TG data for different sizes of Carbon limestone were combined to derive a projcted composite utilization for the size consistancy in the bed of the Babcock and Wilcox 3 ft by 3 ft (0.9 m by 0.9 m) AFBC unit. The average operating parameters of 20 B&W runs (8 ft [2.4 m] bed, 1.5 ft [0.5 m]/s, 50% sulfur removal) using Carbon limestone were used to estimate a molar reaction rate on the TGA (0.1% Ca/min) which corresponds to the average bed rate constant. A weighted average, based on the B&W sorbent feed size distributions, of the sorbent utilization obtained on the TG at 0.1 percent calcium reacting per minute was calculated for TG runs on finely divided size fractions of Carbon limestone at temperatures and gas compositions similar to those in the B&W combustor. The results of this comparison are shown in

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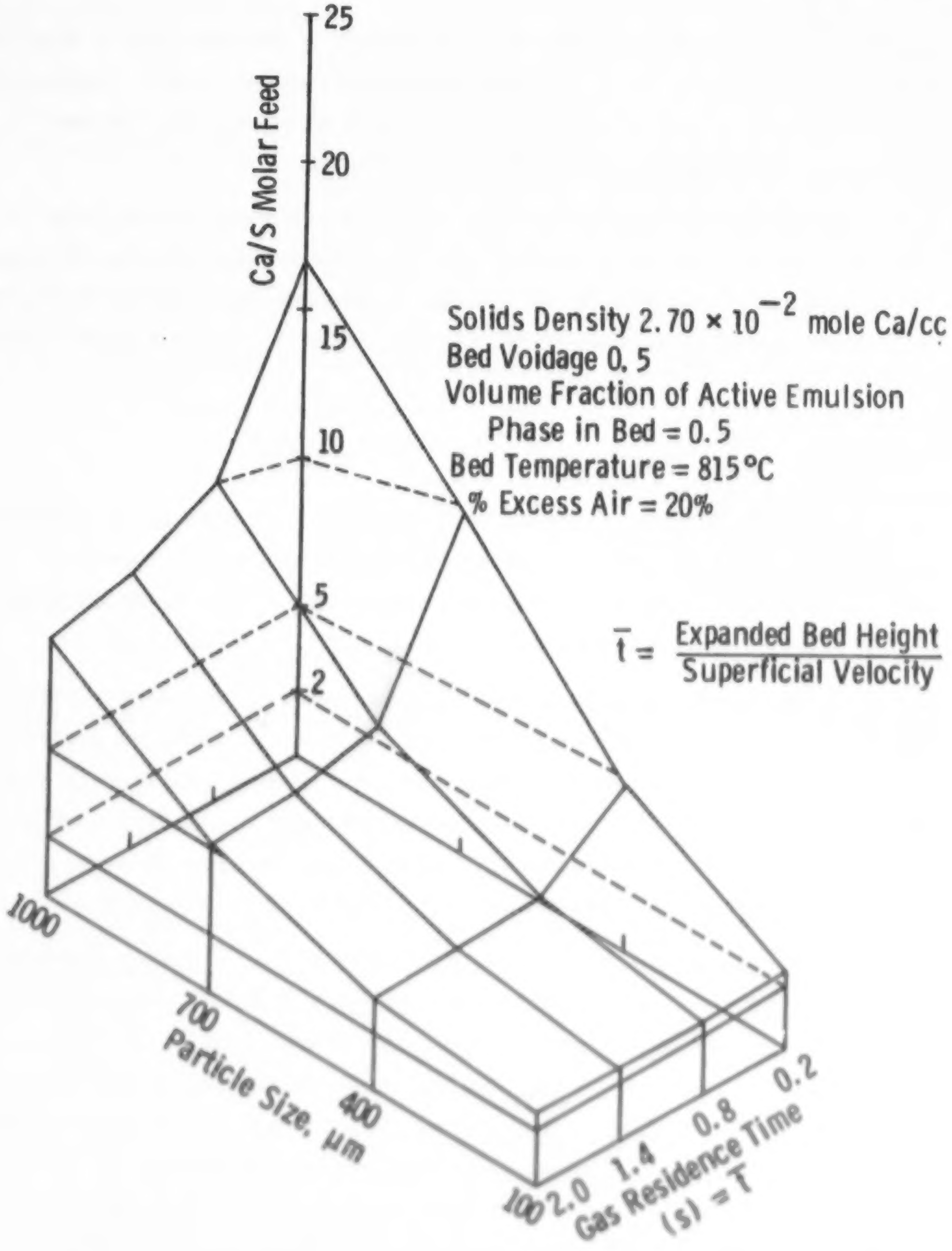

Figure A21 - Ca/S molar feed required to maintain 90% sulfur removal in AFBC with limestone 1359

Curve 690412-A

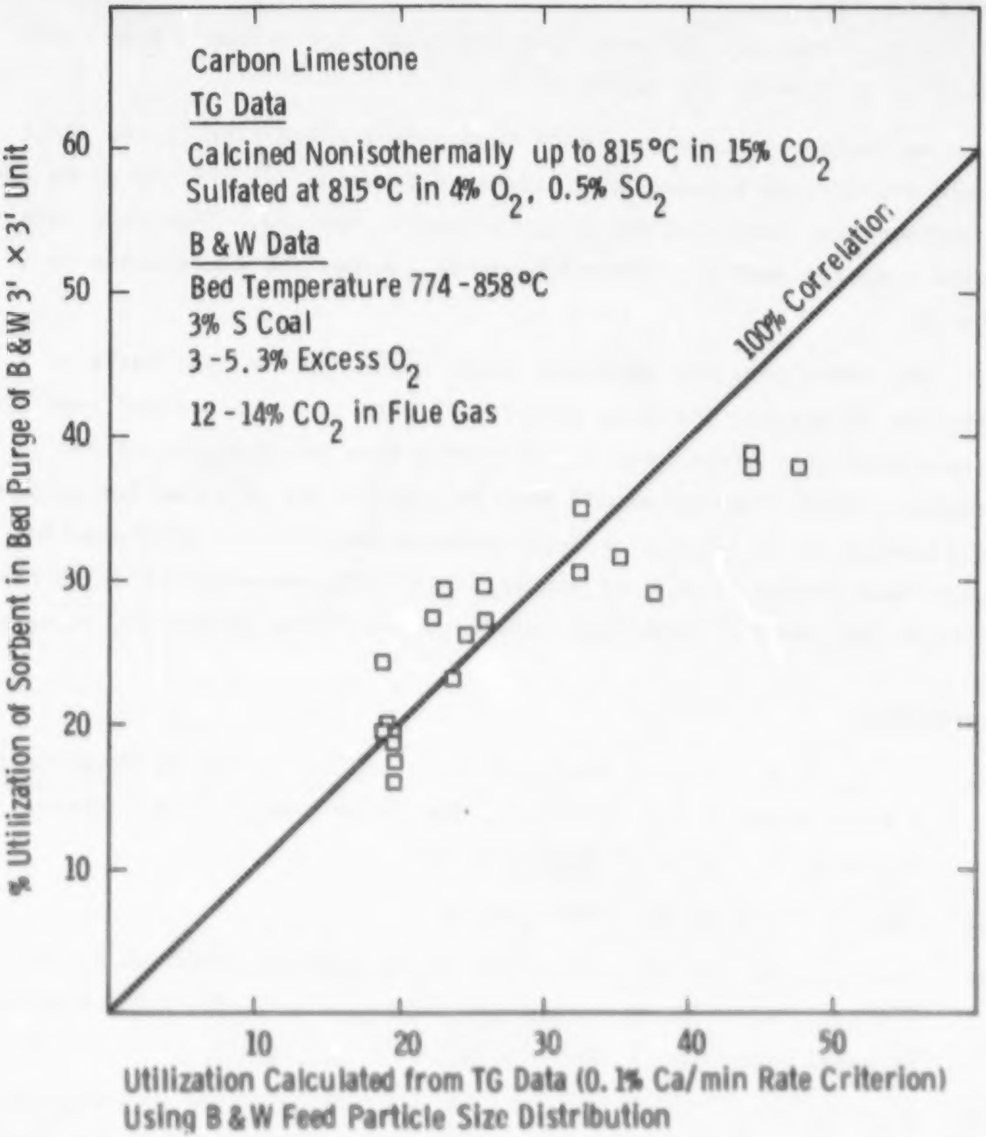

Figure A22 - Comparison of sorbent capacity obtained from Westinghouse TG data with B&W fluid-bed results[(A12)]

Figure A22. Westinghouse feels that the agreement is excellent. This technique to account for particle size distributions is time consuming, however, for many TG experiments are required to cover the full range of particle sizes. A fundamental model that would permit the construction of the TG curves for different particle size ranges from a base curve would be of inestimable value.

As another amendment, a more fundamental correction to the model would result from generation of a mean rate value for the bed given the residence time distribution of the sorbentin the bed. This mean rate would consider particle attrition within the bed and elutriation from the bed.

The third required amendment would adjust the initial rates of reaction to account for mass transfer effects in the fluidized bed. To some extent this effect can be calculated from batch fluidized-bed results. Mass transfer should only be significant in cases for sorbent utilization of 10 percent or less, corresponding to calcium-to-sulfur molar feed ratios of 10/1 or higher. Such high sorbent feed rates lie outside the range of practical interest, except for generative systems.

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APPENDIX B

FBC POWER PLANT DESIGN BASES

In this appendix the various comprehensive conceptual plant designs that have been developed for AFBC and PFBC electric utility power plants are reviewed, and base designs are selected and described. Additional design options are discussed. All of these plants were designed to satisfy current EPA new source performance standards for sulfur oxides (516 ng/J), particulates (43 ng/J), and nitrous oxides (301 ng/J), although, based on current data and Westinghouse projections for sulfur oxide removal, some of the designs will not indeed meet these standards for the specific operating conditions applied (calcium-to-sulfur ratio, bed depth, velocity, sorbent particle size, etc.).

Available cost projections for these designs are compared. The sensitivity of the costs and uncertainties involved in them are discussed, and the base FBC power plant costs are estimated.

CONCEPTUAL DESIGNS FOR AFBC

Comprehensive conceptual designs of AFBC power plants have been reported: by Westinghouse/Foster Wheeler in 1971 to EPA;[B1] by General Electric/Bechtel in 1977 to NASA, ERDA, and NSF (under the Energy Conversion Alternatives study, or "ECAS");[B2] and most recently by Burns and Roe/Combustion Engineering, by Stone & Webster/Foster Wheeler, and by Stone & Webster/Babcock & Wilcox to DOE.[B3] As will be discussed later, the GE/Bechtel designs/costs have been utilized and expanded upon somewhat by TVA.[B7] Table B1 presents a comparison of the key characteristics of these designs. Complete information on all of the designs is not available, but the diversity of the design bases is evident. Combustion Engineering, Babcock & Wilcox, and Foster Wheeler

Table B1

SUMMARY OF CHARACTERISTICS OF AFBC CONCEPTUAL ENGINEERING STUDIES

Dwg. 2616C48

	Westinghouse/Foster Wheeler (EPA) – 1971	General Electric/Bechtel (NASA, ERDA, NSF) – 1977	Burns and Roe/Combustion Engineering (DOE)	Stone & Webster/Foster Wheeler (DOE)	Stone & Webster/Babcock & Wilcox (DOE)
POWER PLANT CONFIGURATION					
Type Plant construction and philosophy	Grassroots; maximum shop fabrication; modular boilers	Grassroots; modular boilers	Maximum shop fabrication of bed cells; grassroots	Minimum land area; Conventional convective heat recovery zone; expansion	Maximum similarity with conventional boilers; expansion
Capacity (MWe)	300 (and 635)	834	553	538	552
Number of boiler modules	4	4	1	1	1
Coal handling	Crushing; drying in coal-fueled fluid-bed dryer with primary air for feeder-distributer	Crushing; drying with waste hot air from spent solids	——	——	——
Sorbent handling	Handling combined with coal handling system Crushed and double-screened sorbent purchased	Handling performed separately from coal handling system; sorbent crushed on site	——	——	——
Spent solids handling	Dry conveying; no capture of solids heat content	Dry conveying; solids air-cooled to 150°C, with air for coal and sorbent drying	Spent solids energy recovery	Spent solids energy recovery	No spent solids energy recovery
Spent solids storage and disposal	Short-term on-site silo, rail disposal with no off-site disposal defined	15 day on-site storage, rail disposal with no off-site disposal defined	——	——	——
Bed-overflow Arrangement	FBC bed-and CBC bed-overflow to spent solids handling	FBC bed-overflow to CBC; CBC bed-overflow to spent solids handling	——	——	——
Cooling tower type	None -- once-through cooling water	Wet mechanical draft-24 cells	——	——	——
FB STEAM GENERATOR DESIGN					
Number of main beds per module	5	6	20 cells	5	5
Arrangement of beds	Vertically stacked; separate functions for each bed	Vertically stacked; separate functions for each bed	Beds on single level	Vertically stacked; separate functions for each bed	2 parallel vertical pairs 5th bed below and between the pairs
Steam circulation/arrangement	Once-through; waterwalls and horizontal tubes	Once-through; waterwalls and horizontal tubes	Drum-type, assisted circulation; waterwalls and horizontal tubes	Once-through; waterwalls and horizontal tubes	Drum-type assisted circulation; waterwalls and horizontal tubes
Steam cycle	Subcritical	Supercritical	Supercritical	Supercritical	Supercritical
Pressure, kPa	16,548	24,133	17,930 (at superheater)	18,070 (at superheater)	18,070 (at superheater)
Superheat Temperature, °C	538	538	541	541	541
Reheat Temperature, °C	538	538	541	541	541
Vessel dimensions					
Height, m	33.2	58.6	42.4	55.5	49.7
Cross-section, m	6.55 (rectangular)	11 × 4 (rectangular)	1784m^2 (rectangular)	366m^2 (rectangular)	685m^2 (rectangular)
Bed area, per bed, m^2	14.5	37.9	30.1 (per cell)	123	158.3
Structural support	Top supported	——	Bottom supported; top supported convection pass	Complete top support	Complete top support
Heat transfer surface					
Configuration	Horizontal	Horizontal	Horizontal	Horizontal	Horizontal
Diameters (cm)	5.08, 3.8	5.08, 3.18	——	——	——
Pitch (cm) (horizontal)	10.2	15.24, 7.6	——	——	——
CONTROL OF CARBON UTILIZATION					
Technique	Carbon burnup cell	Carbon burnup cell	Carbon burnup cell	Carbon burnup cell	Carbon burnup cell
Carbon burnup cell (CBC) design					
Bed area, m^2	~15	~38	100.3	111.5	200.7
Location	Integral bed in main combustor; lowest bed	Integral bed in main combustor, lowest bed	Detached from main beds	Integral bed in main combustor	Detached and to the side of the main combuster
Heat transfer surface	None in bed; in freeboard	Water tubes in freeboard	——	——	——
Fly-ash injection, m^2 per point	Petrocarb; 0.9		1.67	1.0	0.84

Table B1 (Cont'd)

Dwg. 2616C49

	Westinghouse/Foster Wheeler (EPA) — 1971	General Electric/ Bechtel (NASA, ERDA, NSF) — 1977	Burns and Roe/Combustion Engineering (DOE)	Stone & Webster/Foster Wheeler (DOE)	Stone & Webster/Babcock & Wilcox (DOE)
COAL FEEDING					
Technique	Combined dilute-phase pneumatic feeding of coal and sorbent in bed	Combined vibrational feeding of coal and sorbent in bed	Fuller-Kinyon pneumatic in bed feeding; combined	Stoker feeder above bed;	Pneumatic in bed
Bed area per injector point, m^2	0.93	0.8	1.67	10.2	0.84
Feed systems per module	2	7	36 stream splits per coal hopper	15 splits per coal hopper	40 splits per coal hopper
SORBENT FEEDING					
Technique	Combined with coal feeding	Combined with coal feeding	Combined with coal feeding	Gravity feeder above bed	Combined with coal feeding
Bed area per injector point, m^2	Same as coal	Same as coal	Same as coal	60	Same as coal
Feed systems per module	Same as coal	Same as coal	Same as coal	1	Same as coal
BED OVERFLOW					
Technique	Gravity drain	Gravity drain	Gravity drain	Gravity drain	Gravity drain
Bed area per overflow point m^2	14.5	——	10.2	15.3	35.7
PARTICULATE CONTROL					
Equipment	High efficiency multi-clone cyclones; ESP	Cyclones; ESP	Multitube cyclones; baghouse	Multitube cyclones; baghouse	No mechanical collectors; hot ESPs
Arrangement	Cold ESP final cleaning, primary cyclone drains CBC, secondary cyclone drains to disposal	Hot ESP (390°C) for final cleanup	Baghouse for final cleanup	Baghouse for final cleanup	ESPs for primary and final cleaning
COMBUSTOR OPERATING CONDITIONS					
Temperature, °C	871 (1038 CBC)	843 (1093 CBC)	843 (1093 CBC)	843 (1038 CBC)	843 (1093 CBC)
Pressure, kPa	102	102	——	——	——
Excess air, %	10 (70.5 CBC)	20 (30 CBC)	20 (25 CBC)	20 (30 CBC)	15 (50 CBC)
Velocity, m/s	2.4-3.35 (3.3 CBC)	3.05 (2.74 CBC)	3.66 (2.74 CBC)	3.05 (2.44 CBC)	2.44 (2.44 CBC)
Bed depth, m					
Static	——	——	0.61	0.61 (0.70 CBC)	0.46 (0.61 CBC)
Expanded	0.3	1.22	1.22	1.22	1.22
Calcium-to-sulfur ratio	6.0	2.0	2.3	2.1	2.5
Sorbent particle diameter, μm	<6350	<3180	——	——	——
Coal particle diameter, μm	<6350	<12,700	——	——	——
SORBENT					
Type	Limestone	Limestone	Limestone	Limestone	Limestone
Source	1359	——	——	——	——
Size received, μm	<6350 (double screened)	Not crushed or classified	——	——	——
Moisture, %	——	10	——	——	——
COAL					
Type	Bituminous	Bituminous	Western	Eastern	Eastern
Source	Ohio Pgh. No. 8 seam	Illinois No. 6	——	——	——
Composition					
Wt.% sulfur	4.3	3.9	0.89	2.55	2.55
Wt.% ash	8.5	9.6	7.8	10.0	10.0
Wt % moisture	3.3	13.0	28.0	13.6	13.6
Heating value, Lower, J/g	29,073	23,793	18,731	25,774	25,774
PERFORMANCE PROJECTIONS AND ASSUMPTIONS					
SO_2 emission target, ng/J	516	516	516	516	516
Sulfur removal efficiency, %	90 projected	85 projected	43.0 required	73.9 required	73.9 required
Particulate target, ng/J	43; 0.01 grains per scf projected	43	43	43	43
Availability Target, %	——	90	——	——	——
Boiler efficiency, %	86.1	87.9	84.2	85.93	85.83
Combustion efficiency, %	87 (in main beds), 85 (CBC)	98.4	——	——	——
Plant efficiency, %	35.7	35.6	34.2	34.1	34.6
Auxiliary power, % of gross	——	7.3	——	——	——
Stack temperature, °C	135	121	——	——	——
CBC heat duty, % of total	11.6	1[illegible].0	——	——	——

Table B1 (Cont'd)

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	Westinghouse/Foster Wheeler (EPA) – 1971	General Electric/ Bechtel (ERDA) – 1977	Burns and Roe/Combustion Engineering (DOE)	Stone & Webster/Foster Wheeler (DOE)	Stone & Webster/Babcock & Wilcox (DOE)
Air Supply					
FD fan pressure, kPa	113 (preheat to 390°C)	118 (no preheat)	—	—	—
ID fan pressure, kPa	none	2.1	—	—	—
System pressure drop, kPa	11.8	18.3	—	—	—
Heat Transfer coefficients, J/sm^2 °C					
In bed	283	226	256	227	256-284
Above bed	226 (in splash region)	74	—	—	—
Particulate loading basis					
Ash, % of ash input	100	—	100	100	75
Sorbent, % of fresh feed	3	—	—	—	—

each are performing 200 MWe fluidized-bed boiler demonstration unit designs for TVA, from which commercial plant designs are being scaled, but these designs were not available at the time this report was prepared.

Major differences between the conceptual designs exist in: the type of plant construction (grass roots or expansion); the design philosophy (boiler modularity, maximum shop fabrication, minimum land area, maximum similarity with conventional boilers - especially in the convective pass); type of cooling tower (once-through cooling water or mechanical draft tower, type of steam circulation, coal and sorbent handling [combined or separate handling]); solids feeding technique (in bed or above bed, bed area per feed point, combined or separate coal and sorbent feeding); particulate control equipment (cyclones and electrostatic precipitator [ESP], cyclones and hot ESP, cyclones and baghouse, hot ESP with no mechanical collectors). Major differences in the fluid-bed combustor operating conditions are found in the superficial velocity (2.4 to 3.66 m/s) and the excess air levels (10 to 20% for the main beds and 25 to 70.5% for the carbon burnup cell). Significant differences in the coal properties for each design also exist. The most significant differences in plant performance are in the projected boiler efficiency (84.2 to 87.9%) and the projected plant efficiency (34.1 to 35.8%). These would be influenced by the auxiliary power consumption, the assumed carbon combustion efficiency and the recovery of energy from the hot spent solids.

For every design category in Table B1 there are a multitude of options in addition to those selected in the five conceptual design studies listed. For the dense-phase fluidization concepts currently being developed for AFBC with once-through sorbent operation, the most important additional options not considered in these conceptual designs are related to the combustion efficiency of carbon (carbon burnup cell, fines recycle to main beds, low carbon elutriation by low gas velocity, high excess air, high bed temperature, coal particle size), and the disposal of spent sorbent (landfill, ocean dumping, long-term storage, processing for environmental acceptability or for utilization as building material, agricultural uses, etc.).

Several alternative concepts for the FBC sulfur removal system are also of potential interest but are not considered here. For example,

- Sorbent regeneration
  - reductive decomposition
  - two-step sulfide generation, H_2 generation
  - sulfide-sulfate solid-solid reaction
- Precalcination of once-through sorbents for improved reactivity and utilization
- Additives for improved limestone/dolomite utilization
- Reconstitution of sorbent fines by agglomeration
- Alternative metal oxide sorbents.

These and other sulfur removal system concepts are under study but have not been considered in the first-generation AFBC design studies. Their impact on FBC cost and performance could ultimately be significant.

Cost projections for the Westinghouse[B1] and General Electric[B2] conceptual designs are the major AFBC designs available at the time of this writing. To enable comparison the Westinghouse cost estimates were converted, as required, to the costing assumptions used in the body of this report (see Table 1, page 11); the GE estimates, having been prepared under ECAS, were already based upon the assumptions in Table 1. The 1971 Westinghouse costs have been escalated to mid-1975 dollars using the Chemical Engineering Cost Index, which is very approximate in nature. The costs are \$571/kW for the Westinghouse design (converted to the Table 1 assumptions) and \$632/kW for the General Electric design. This is only a 10% difference in capital investment, even with the differences in plant design basis (crushed sorbent purchased, no spent solids heat recovery, no cooling tower in Westinghouse AFBC power plant design) and the approximate nature of the above escalation of the Westinghouse cost. A probable reason for the fairly close agreement is that the AFBC power plant is at least 75 percent conventional equipment with respect to capital investment. For example, the General Electric AFBC plant costs break down to 16.1 percent for the fluid-bed combustor, 1.6 percent for solids feeding, 3.0 percent for spent solids cooling, and 4.1 percent for particulate control. These four systems may be considered to be nonconventional, although they all represent systems comprised of largely conventional components. Because of this, AFBC power plant capital investment projections by different plant designers applying different AFBC design concepts are expected to be reasonably close together (±15%) if the costs are placed on the same basis and the power plant scopes are identical.

The sensitivity of the capital cost of the fluid-bed boiler and the AFBC power plant to changes in design and operating conditions is expected to be small. The fluid-bed boiler can be minimized in cross-sectional area by operating with high superficial velocities, using large sorbent particle diameters and low excess air rates. This reduction in cross-section will require a simultaneous increase in bed depth in order to provide sufficient volume for the immersed heat transfer surface and sufficient gas-solid contacting time to meet combustion efficiency and sulfur removal targets. The increase in bed depth, and the resulting increase in freeboard height, will counteract, to a large extent, any cost reduction due to decreased vessel cross-section. Ultimately, changes in design and operating conditions that increase the AFBC power plant efficiency, such as increased carbon combustion or reduced sorbent consumption by increasing bed depth or reducing superficial velocity, will have a greater impact on the cost of electricity than will the combustor cross-sectional area.

Cost and design uncertainties for AFBC exist mainly in the areas of performance and reliability of the solids feeding system, the fluid-bed boiler, and the particulate control system. Long-term erosion, corrosion, plugging and deposition phenomena (occurring to tubes, distributor plates, solids feed lines, cyclones, etc.) could affect the plant reliability but may not ultimately have a great cost impact if these phenomena can be controlled by proper selection of operating conditions and hardware design. The number of solids feed points required to promote uniform bed conditions, high carbon combustion efficiencies, reliable boiler operation, and the like is an example of an uncertainty in AFBC that should be resolved through early large-scale operation and will, it is hoped, have little cost impact on AFBC. Innovations such as above-bed feeding conceivably could simplify the solids feeding system and improve reliability, though performance problems in sulfur removal and carbon utilization may result.

TVA[B7] has applied cost uncertainty factors (an add-on multiplier) of 20 percent to the General Electric costs for spent-bed coolers and cyclones, coal and limestone blending and feeding, hot gas cleanup, air supply and fines injection, and the AFB module (including tower components, controls and ducts). TVA claims that, on the basis of its experience, such factors account for potential increases in the cost of these undemonstrated components. The TVA uncertainty factors increase the General Electric base investment by only [illegible].7 percent, an amount that is probably not significant, based on the expected accuracy of conceptual design cost estimates. TVA also suggested other modifications to the AFBC power plant design, such as in the area of solid residue disposal (see Appendix C).

CONCEPTUAL DESIGNS FOR PFBC

Westinghouse/Foster Wheeler,[B1] Westinghouse[B4] (update of Reference B1), Westinghouse/C. R. Main (under ECAS),[B5] and General Electric/Bechtel (under ECAS)[B2] have reported on comprehensive conceptual designs for PFBC. These designs are characterized and compared in Table B2. The GE/Bechtel design has been utilized and somewhat expanded upon by TVA.[B7] The concept dealt with in these studies is the fluidized-bed boiler (with steam tubes in the bed) with an open-cycle gas turbine expanding the exhaust gas. Many other cycles are possible, ranging from the adiabatic combustor to air-cooled FBC cycles to various closed gas-turbine cycles, along with several other fluid-bed combustor concepts, such as recirculating FBC concepts. Other conceptual designs for PFBC power plants are being developed by Burns & Roe/United Technologies/Babcock & Wilcox, Curtiss-Wright/Dorr-Oliver/Stone & Webster, General Electric/Foster Wheeler under DOE sponsorship and privately by American Electric Power.

Many similarities exist in the four PFBC designs summarized in Table B2. The major differences are found in the solids feeding system, the particulate control equipment design, and the combustor operating conditions. Even though Westinghouse/C. T. Main and General Electric/Bechtel use widely different gas turbine inlet temperatures, 960 and 871°C, respectively, their projected power plant efficiencies are nearly identical at 39.0 and 39.2%, respectively. Westinghouse and General Electric also scaled their designs to a 904°C turbine temperature. The General Electric/Bechtel design used coarse sorbent particle sizes and high fluidization velocities relative to the Westinghouse/C. T. Main design. It is not surprising that the systems that differ most between the Westinghouse ECAS and General Electric ECAS designs, the solids feeding system and the particulate control system, represent the areas of greatest design uncertainty. As with the AFBC design, many options exist with the PFBC design. Several methods for controlling carbon utilization are possible, such as carbon burnup cells, fines recycle to the main combustor beds, operation with high excess air and/or low fluidization velocities, operation with high combustor temperatures, and control of coal feed-size.

Solids feeding could be performed by either in-bed or above-bed techniques. As in AFBC, the number of feed points per unit of bed area, the impact of above-bed feeding, and the mechanical reliability of available feeding systems are not understood. Current conceptual designs use a number of feed points which, the designers feel, have been sufficiently demonstrated in development units to provide a conservative basis.

Table B2

SUMMARY OF CHARACTERISTICS OF PFBC CONCEPTUAL ENGINEERING STUDIES

Dwg. 2614C50

	Westinghouse/Foster Wheeler (EPA) – 1971	Westinghouse Update (EPA) – 1975	Westinghouse/C.T. Main (NASA, ERDA, NSF) – 1977	General Electric/Bechtel (NASA, ERDA, NSF) – 1977
POWER PLANT CONFIGURATION				
Type of plant construction and philosophy	Grassroots; modular with maximum shop fabrication	Grassroots; modular construction	Grassroots; modular construction	Grassroots; modular construction
Capacity (MWe)	318 (and 635)	635	619	904
Number of boiler modules	4	4	4	4
Number of gas turbines	2	2	2	4
Gas turbine compression ratio	10:1	10:1	10:1	10:1
Coal handling	Crushing; drying in coal-fueled fluid-bed dryer	Crushing; drying in coal-fueled fluid-bed dryer	Combined coal and sorbent crushing and drying to 3% moisture in hot-air dryer	Separate crushing and drying to 3% moisture in hot-air dryer
Sorbent handling	No crushing (purchased sized); drying combined with coal handling	No crushing (purchased sized); drying combined with coal handling	Combined with coal	Separate crushing and drying to 3% moisture in hot air dryer
Spent solids handling	Dry conveying with no energy recovery	Dry conveying with no energy recovery	Cooling with energy recovery	Cooling with energy recovery
Spent solids storage and disposal	Short-term on-site silo; rail disposal; no off-site storage	Short-term on-site silo; rail disposal; no off-site storage	15 day on-site storage; no off-site storage	15 day on-site storage; no off-site storage
Bed-overflow arrangement	FBC bed and CBC bed overflow to spent solids handling	FBC bed and CBC bed overflow to spent solids handling	FBC bed and CBC bed overflow to spent solids handling	FBC bed overflow to CBC; CBC bed overflow to spent solids handling
Cooling tower type	None -- once-through cooling water	None -- once-through cooling water	Mechanical draft evaporator	Wet mechanical draft evaporator
FB COMBUSTOR DESIGN				
Number of main beds per module	4	4	4	6
Arrangement of beds	Vertically stacked; separate functions for each bed	Vertically stacked; separate functions for each bed	Vertically stacked; separate functions for each bed	Vertically stacked; separate functions for each bed
Steam circulation/arrangement	Once-through Waterwalls and horizontal tubes	Once-through Waterwalls and horizontal tubes	Once-through Waterwalls and horizontal tubes	Once-through; Waterwalls and horizontal tubes
Steam cycle	Subcritical	Subcritical	Supercritical	Supercritical;
Pressure, kPa	16,548	16,546	24,133	24,133
Superheat temperature, °C	538	538	538	538
Reheat temperature, °C	538	538	538	538
Vessel dimensions				
Height, m	37	37	28.3	36.6
Cross-section, m	3.7 (cylindrical)	5.2 (cylindrical)	6.1 (cylindrical)	4.0 (cylindrical)
Bed area, per bed, m^2	3.25	6.5	8.5	4.0
Structural support	Top supported	Top supported		
Heat transfer surface				
Configuration	Horizontal	Horizontal	Horizontal	Horizontal
Diameter, cm	5.1, 3.9	5.1, 3.9	5.1, 3.8	4.4, 3.2
Horizontal pitch, cm	10.2	10.2	8.9	11.4, 7.6
CONTROL OF CARBON UTILIZATION				
Technique	Carbon burnup cell	Carbon burnup cell	Carbon burnup cell	Carbon burnup cell
Carbon burnup cell (CBC) design				
Bed area, m^2	1.3	2.6	3.0	2.8
Location	Integral bed in combustor	Integral bed in combustor	Integral bed in conbustor	Integral bed in combuster
Heat transfer surface	Waterwalls	Water walls	Waterwalls and in-bed surface	No surface
Fly ash injection, m^2 per point	0.33	0.66	~1.0	——
COAL FEEDING				
Technique	Petrocarb; combined coal and sorbent feeding	Petrocarb; combined coal and sorbent feeding	Petrocarb; combined coal and sorbent feeding	Petrocarb
Bed area per injector point, m^2	0.8	0.8	1.07 – 2.14	0.55
Feed systems per module	4	4	4	6

Table B2 (Cont'd)

Dwg. 2616C51

	Westinghouse/Foster Wheeler (EPA) – 1971	Westinghouse Update (EPA) – 1975	Westinghouse/C.T. Main (NASA, ERDA, NSF) - 1977	General Electric/Bechtel (NASA, ERDA, NSF) - 1977
SORBENT FEEDING				
Technique	Combined with coal	Combined with coal	Combined with coal	Petrocarb
Bed area per injector point, m^2	Same as coal	Same as coal	Same as coal	0.55
Feed systems per module	Same as coal	Same as coal	Same as coal	6
BED OVERFLOW				
Technique	Gravity flow	Gravity flow	Gravity flow	Gravity flow
Bed area per overflow point, m^2	3.25	6.5	8.5	4.0
PARTICULATE CONTROL				
Equipment	Two stages of cyclones	Two stages of cyclones plus final-stage granular bed filter (GBF)	Two stages of cyclones plus Ducon GBF	Two stages of cyclones plus Ducon GBF
Arrangement	Primary (1 Ducon per module) drains to CBC; secondary (2 Aerodynes) drains to disposal	Primary drains to CBC; secondary and GBF (4 per module) drain to disposal	Primary cyclone drains to CBC; secondary and GBF drain to disposal	Primary cyclone drains to CBC; secondary and GBF drain to disposal
COMBUSTOR OPERATING CONDITIONS				
Temperature, °C	954 (1093 CBC)	954 (1093 CBC)	1010 (1010 CBC)	899 (1093 CBC)
Pressure, kPa	1013	1013	1013	1013
Excess air, %	10 (79 CBC)	17.5 (79 CBC)	6 (36 CBC)	20 (30 CBC)
Velocity, m/s	1.7–2.7 (1.74 CBC)	1.7–2.7 (1.74 CBC)	1.5–1.92 (0.91 CBC)	2.1 (1.7 CBC)
Bed depth (expanded), m	3.7	3.7	2.1–2.7 (1.5 CBC)	2.4 (1.8 CBC)
Calcium-to-sulfur ratio	6.0	2.0	1.5	2.0
Sorbent particle diameter, um	2500 (average)	1000 (average)	<3170; 800 (average)	<6350
Coal particle diameter, um	<6350	<6350	<6350; 1400 (average)	<6350; 1400 (average)
SORBENT				
Type	Dolomite	Dolomite	Dolomite	Dolomite
Source	1337	1337	——	——
Size received, um	<6350; 2500 (average)	——	——	——
Moisture, %	——	——	10	10
COAL				
Type	Bituminous	Bituminous	Bituminous	Bituminous
Source	Ohio Pgh. No. 8 Seam		Illinois No. 6	Illinois No. 6
Composition				
Wt% sulfur	4.3	3	3.9	3.9
Wt% ash	8.5	——	9.6	9.6
Wt% moisture	3.3	——	13.0	13.0
Heating value, lower, J/g	29,073	29,073	23,793	23,793
PERFORMANCE PROJECTIONS AND ASSUMPTIONS				
SO_2 emission target, ng/J	516	516	516	516
Sulfur removal efficiency, %	90 projected	85 projected	84 required	84 required
Particulate target, ng/J	43	43	43	43
Availability target, %	——	——	90	90
Boiler efficiency, %	88.6	92.0	——	——
Combustion efficiency, %	94 (in main beds), 90 (in CBC)	——	95 (in main beds), 90 (in CBC)	98.5
Plant efficiency, %	38.1	37.7	39.0	39.2
Power distribution, % gas turbine	17.4	——	20	22
Auxiliary power, % of gross	2.8	——	3.6	4.2
Combustor exit temperature, °C	899	——	982	871
Gas turbine inlet temperature, °C	871	——	960	871
Stack temperature, °C	135	——	135	149
CBC heat duty, % of total	6.0	——	4.5	4
Air Supply				
Pressure, kPa	1005	1005	1005	1003
Pressure drop, kPa	120	120	80.7	51.7
Heat Transfer Coefficients, J(sm^2°C)$^{-1}$				
In bed	283		272	283
Above bed	170-220 (splash region)		85	96

Table B2 (Cont'd)

Dwg. [illegible]

	Westinghouse/Foster Wheeler (EPA) – 1971	Westinghouse Update (EPA) – 1975	Westinghouse/C.T. Main (ERDA) – 1977	General Electric/Bechtel (ERDA) – 1977
Particulate Loading Basis				
Ash, % of ash input	100	100	37, 460 ppm at combuster exit; 750 ppm to GBF	43, 870 ppm at combuster exit; 1790 ppm to GBF
Sorbent, % of fresh feed	5	5		
Granular Bed Filter Basis	no granular bed filter system	——		
Face velocity, m/min.			75	39
Number of filters per boiler module	——	——	4	16

The conceptual designs reviewed used granular bed filters for final-stage particulate control. The critical specification, such as face-velocity, number of granular bed modules, and the granular bed filter reliability, is currently not known, though test programs are proceeding that deal with these questions. In addition to granular bed filters, other high-temperature filters, such as ceramic filters and even high-temperature electrostatic precipitators, are being proposed for final-stage particulate control. Low-temperature devices such as electrostatic precipitators or baghouses with recuperative heat exchange systems are probably too complex and thermally inefficient to be considered for PFBC.

In contrast with AFBC, a much greater range of operating conditions may be applicable to PFBC power generation systems than are included in the conceptual designs reviewed. Higher gas-turbine pressures, such as in the Americal Electric Power PFBC concept, alternative combustor temperature ranges, and higher excess air levels, such as in the adiabatic combustor concepts, may lead to quite different PFBC power plant performance and economics.

Alternative sorbent system concepts are under development for PFBC, in small-scale efforts that may improve PFBC economics and environmental impact by reducing sorbent consumption. These include sorbent regeneration schemes, precalcination of sorbents, additives that improve sorbent utilization, and alternative sorbents.

The four available cost projections for the PFBC conceptual designs have been converted to an identical set of cost ground rules (Table 1) for comparison. The Westinghouse and GE estimates under ECAS were already prepared according to the assumptions in Table 1 and needed no conversion. Significant equipment and performance assumption differences account for the differences in the power plant investments:

Investment	Westinghouse[B1] 1971 (EPA)	Westinghouse[B4] 1975 (EPA)	Westinghouse[B5] 1977 NASA/ ERDA/NSF	General Electric[B2] 1977 NASA/ ERDA/NSF
$/kW	527	580	610	723

For example, in the Westinghouse ECAS and General Electric ECAS studies the 18.5 percent difference in investment can be attributed mainly to differences in cost estimates for solids feeding and particulate control. The change in energy cost due to this investment difference would be about 10 percent, and the significance of a 10 percent difference between conceptual designs would generally be considered to be low.

Differences in the 1971 and 1975 Westinghouse cost estimates and the Westinghouse ECAS cost are attributed mainly to plant basis and equipment selection. For example, Table B2 indicates that the Westinghouse 1971 and 1975 conceptual PFBC design purchased double-screened sorbent rather than crushing and screening on site and used once-through cooling water rather than a cooling tower as in the 1977 Westinghouse conceptual design. The 1971 Westinghouse PFBC design used multiple stages of high-efficiency cyclones for particulate control, and the 1975 Westinghouse update added a final-stage granular bed filter. In general, then, these costs are consistant and represent the only four comprehensive PFBC conceptual designs currently available.

The breakdown of the capital investment of PFBC indicates, as with AFBC, the power plant is largely conventional equipment (70-80% with respect to investment). The pressurized fluidized-bed boiler represents only about 6 to 8 percent of the total plant investment, so efforts to reduce its cost further will probably not be effective in reducing the cost of PFBC power generation.

In addition, in 1973[B6] Westinghouse carried out a detailed cost sensitivity analysis of PFBC. This study showed that the plant capital investment would vary only over a range of about 8 percent of the base investment because of conceivable changes in the pressurized boiler operating conditions (other than the calcium-to-sulfur ratio), such as bed temperature, velocity, bed depth, excess air, and pressure; changes in particulate carry-over from the boilers; changes in the heat transfer surface configuration, tube materials, and heat transfer coefficient; and changes in the gas-turbine inlet temperature and steam superheat temperature. The cost of the fluid-bed boiler itself was especially insensitive to design and operating changes. This is a very important conclusion with respect to efforts to reduce the fluid-bed combustor cost by reducing its diameter with higher operating velocities. Since design and operating changes produce little impact on cost, the performance of the combustor (e.g., combustion efficiency, particle elutriation, calcium utilization) should be maximized in order to move in the direction of optimum PFBC economics.

Uncertainties for PFBC rest mainly with the solids feeding and particulate control systems and their impact on PFBC power plant reliability. Additional uncertainties, as with AFBC, are based on general erosion, corrosion, deposition, and agglomeration phenomena that could occur and could reduce PFBC plant performance and reliability. This latter class of uncertainties may be resolved through proper materials selection, operating conditions selection, and minor engineering modifications. TVA[B7] reviewed the General Electric ECAS PFBC conceptual design and additional costs to components that they considered unconventional. These cost factors, TVA claims, based on their experience with developmental processes, reflect the uncertainties related to these components. They applied factors of 20 percent to increase the investments associated with coal injection, sorbent injection and particulate control, and factors of 40 percent to the investment for spent bed material handling components and the pressurized fluid-bed boiler modules. These cost increases resulted in an increased power plant investment of about 12 percent over the General Electric base cost. As with the TVA review of PFBC, the cost allowances applied by TVA for PFBC do not account for the possibility of cost reductions with the components they considered to be uncertain. The TVA assessment of the GE/Betchel design includes considerations of spent bed material disposal; this aspect of the TVA assessment is more meaningful than are the cost allowances for residue disposal included in other conceptual designs.

SELECTED FBC POWER PLANT BASES

FBC power plant bases to be used in this study for AFBC and PFBC have been selected from the conceptual engineering studies characterized in Tables B1 and B2. The characteristics of the selected base FBC power plants are summarized in Table B3. The AFBC and PFBC power plants were taken mainly from the General Electric[B2] ECAS conceptual designs. General Electric designs provided the basic framework because of the subsequent conventional power plant design that was performed by GE under the same framework and because of the review by TVA.[B7] As discussed below, some major components of the GE design have been adjusted on the basis of the Westinghouse PFBC design under ECAS[B5] and previous Westinghouse work.[B1,B4].

The AFBC base plant design applied is essentially taken directly from the General Electric ECAS study, but TVA has added off-site storage in a lined pond to the plant, and Westinghouse has substituted a baghouse system for final-stage particulate control for the original electrostatic precipitator.

The AFBC power plant cost is categorized into seven groups: steam generators; turbine generators; process mechanical equipment; electrical; civil, and structural; process piping and instrumentation; yardwork and miscellaneous; and costs are reported as major component costs and balance-of-plant costs. From these seven groups, changes were made only to the steam generator and the process mechanical equipment categories for the AFBC base plant. In the process mechanical equipment category, the cost of off-site spent solids storage was added to the solids handling equipment, shown in Table B4 and based on TVA[B7] estimates for off-site storage in a clay-lined, diked impoundment. The steam generator category includes hot gas cleanup equipment, air supply equipment, and the fluid-bed boiler tower components. The hot gas cleanup category is modified by substitution of a baghouse filter system for the original electrostatic precipitators. These systems are assumed to be identical in investment cost for the purposes of this study, as is shown in Table B5. No changes are made in the original tower component costs, Table B6. Also, no changes are made in the balance-of-plant costs (direct labor, indirect field labor, and material(s). Table B7 summarizes the AFBC base plant cost, comparing it with the original GE plant estimate. It must be stressed that the base design conditions (Table B3) and costs (Table B7) have no particular significance -- they do not represent optimum conditions or a representative cost, but are developed only as a means for scaling the plant costs to parametric operating conditions (Appendix C).

The PFBC base plant design is the GE ECAS estimate modified to substitute the Westinghouse ECAS designs for coal and sorbent handling, solids feeding and final-stage granular bed filter particulate control designs/cost estimates in place of the GE estimates for those items. The PFBC combustor operating temperature is also increased from the General Electric base-value of 899°C to 954°C, without changing any of the General Electric base costs, and off-site storage in a lined pond has been added by TVA.

As with AFBC, seven cost categories were used to describe the PFBC power plant cost and only two of these were modified from the GE ECAS costs -- the process mechanical equipment category and the PFB steam generator category. The major cost differences between the GE and Westinghouse PFBC designs are summarized in Table B8. The major items, which can be explained by differences in basic design philosophy, are coal and dolomite injection and the granular bed filter.

Table B9 lists the major equipment costs in the solids handling equipment segment of the process mechanical equipment category, with both the GE original cost estimates and the base PFBC plant cost estimates being listed. Tables B2 and B10 summarize the differences in design bases for coal and sorbent injection equipment adopted in the two designs, the Westinghouse basis being applied for the base design. In general, the GE basis is much more conservative than the Westinghouse.

Table B3

Dwn. 1694B61

CHARACTERISTICS OF SELECTED BASE FBC DESIGNS

	AFBC Base Design	PFBC Base Design
POWER PLANT CONFIGURATION		
Philosophy	Modular construction	Modular construction
Type of construction	Grassroots	Grassroots
Capacity	814^a	904^a
Number of boiler modules	4	4
Number of gas turbines	N.A.	4
Gas turbine compression	N.A.	10:1
Coal handling	Crushing; drying with waste hot air from spent solids energy recovery	Combined coal and sorbent^c crushing and drying to 3% moisture in hot-air dryer
Sorbent handling	Handling separate from coal; sorbent crushing and drying on site	Combined with coal
Spent solids handling	Dry conveying; solids air cooled to 150°C, with air for coal and sorbent drying	Cooling with energy recovery for coal and sorbent drying
Spent solids storage	15 day on-site storage, off-site hauling to lined pond	15 day on-site storage, off-site hauling to lined pond
Bed-overflow arrangement	FBC main bed overflow to CBC; CBC overflow to spent solids handling	FBC main bed overflow to CBC; CBC overflow to spend solids handling
Cooling tower type	Wet mechanical draft	Wet mechanical draft
FB COMBUSTOR DESIGN		
Number of main beds per module	6	6
Arrangement of beds	Vertically stacked; separate functions for each bed	Vertically stacked; separate functions for each bed
Steam circulation/ arrangement	Once-through; waterwalls and horizontal tubes	Once-through; waterwalls and horizontal tubes
Steam cycle	Supercritical	Supercritical
Pressure, kPa	24,133	24,133
Superheat temperature, °C	538	538
Reheat temperature, °C	538	538
Vessel dimensions		
Height, m	58.6^b	36.6^b
Cross-section, m	11×4^b (rectangular)	4.0^b (cylindrical)
Bed area, per bed, m^2	37.9^b	4.0^b
Heat transfer surface		
configuration	Horizontal	Horizontal
diameter, cm	5.08, 3.18	4.4, 3.2
horizontal pitch, cm	15.24, 7.6	11.4, 7.6

Table B3 (Cont'd)

Dwg. 1694B62

	AFB Base Design	PFBC Base Design
CONTROL OF CARBON UTILIZATION		
Technique	Carbon burnup cell	Carbon burnup cell
Carbon burnup cell design		
Bed area, m^2	~38[b]	2.8[b]
Location	Integral bed in combustor	Integral bed in combustor
Heat transfer surface	Tubes in freeboard	No surface
COAL FEEDING		
Technique	Combined vibrational, in-bed, feeding of coal and sorbent	Petrocarb; combined coal and sorbent feeding[c]
Bed area per injector, m^2	0.8	1.07
Feed systems per module	7	4
SORBENT FEEDING		
Technique	Combined with coal feeding	Combined with coal[c]
Bed area per injector, m^2	Same as coal	Same as coal
Feed systems per module	Same as coal	Same as coal
BED OVERFLOW		
Technique	Gravity flow	Gravity flow
Bed area per overflow point, m^2	——	4.0[b]
PARTICULATE CONTROL		
Equipment	2 cyclone stages and baghouse system	Two stages of cyclone plus Ducon Granular Bed Filters (GBF)[c]
Arrangement	Primary cyclone drains to CBC; secondary cyclone and baghouse drain to disposal	Primary cyclone drains to CBC; secondary and GBF drain to disposal
COMBUSTOR OPERATING CONDITIONS		
Temperature, °C	843 (1093 CBC)	954 (1093 CBC)
Pressure, kPa	102	1013
Excess air, %	20 (30 CBC)	20 (30 CBC)
Velocity, m/s	3.05 (2.74 CBC)[d]	2.1 (1.7 CBC)[d]
Bed depth (expanded), m	1.22	2.4 (1.8 CBC)[d]
Calcium-to-sulfur ratio	2.0[d,e]	2.0[d,e]
Sorbent particle diameter, μm	<3180[d]	<6350[d]
Coal particle diameter, μm	<12,700	<6350
SORBENT		
Type	Limestone	Dolomite
Source	e	e
Size received, um	No preclassification	No preclassification
Moisture, %	10	10

Table B3 (Cont'd)

Dwg. 1694B63

	AFBC Base Design	PFBC Base Design
COAL		
Type	Bituminous[d]	Bituminous[d]
Source	Illinois No. 6[d]	Illinois No. 6[d]
Composition		
Wt % sulfur	3.9[d]	3.9[d]
Wt % ash	9.6[d]	9.6[d]
Wt % moisture	13.0	13.0
Heating value, lower, J/g	23,793	23,793
PERFORMANCE PROJECTIONS AND ASSUMPTIONS		
SO$_2$ emission target, ng/J	516[d]	516[d]
Sulfur removal efficiency, %	84 required[e]	84 required[e]
Particulate target, ng/J	43[d]	43[d]
Availability target, %	90	90
Boiler efficiency, %	87.9	——
Combustion efficiency, %	98.4[a]	98.5[a]
Plant efficiency, %	35.8[a]	39.2[a]
Auxiliary power, % of gross	7.3[a]	4.2[a]
Combustor exit temperature, °C	N.A.	926
Gas turbine inlet temperature, °C	N.A.	904
Stack temperature, °C	121	149
CBC heat duty, % of total	10.0[b]	4[b]
Power distribution, % gas turbine	N.A.	22
Air supply		
Supply pressure, kPa	118	1003
ID fan pressure, kPa	2.1	N.A.
Pressure drop, kPa	18.3	51.7[b]
Heat transfer coefficient J (sm^2 °C)$^{-1}$		
In bed	226	283
Above bed	74	96
Particulate loading Basis		
Ash, % of ash input	100	100
Sorbent, % of fresh feed	(See Appendix D)	(See Appendix D)
Granular bed filter basis	N.A.	
Face velocity, m/min	——	75[c]
Number per boiler module	——	4[c]

a. This represents the reference value for this quantity. Parametric values for this quantity have also been determined as a function of the study parameters.

b. This represents the reference value. The value is modified in the parametric studies, but is not specifically reported.

c. For further design details see Reference B5.

d. This is a major parameter in this study and the value shown is the reference value.

e. Specific projections for this quantity are reported. See Appendix A.

N. A. = Not applicable

Table B4

SOLIDS HANDLING MAJOR EQUIPMENT FOR AFBC

Subsystems	GE Cost Estimate, 1975 M$	Number	Original GE Estimate, M$	Estimate Used in Base Plant Cost, M$
COAL DRYING & CRUSHING				
1 - Dryer System-189 TPH	0.910			
1 - Coal Crusher & 2 Screens	0.159			
1 - Distribution Box	0.031			
2 - Vibrating Feeders @ 94-1/2 TPH	0.012			
2 - Surge Bins @ 7600 ft^3	0.060			
2 - Bin Activators	0.021			
2 - Weigh Belt Feeders @ 94-1/2 TPH	0.053			
Section Subtotal =	1.246	3	3.74	3.74
LIMESTONE DRYING & CRUSHING				
1 - Dryer System @ 48.6 TPH	0.359			
1 - Crusher & 2 Screens	0.033			
1 - Distribution Box	0.011			
2 - Vibrating Feeders @ 24.3 TPH	0.008			
2 - Surge Bins	0.022			
4 - Bin Activators	0.026			
2 - Air Lock Valves	0.006			
2 - Vibrating Feeders	0.008			
2 - Weighbelt Feeders @ 24.3 TPH	0.018			
Section Subtotal =	0.541	3	1.62	1.62
COAL & LIMESTONE BLENDING & FEEDING				
2 - Blenders @ 115 TPH	0.130			
14 - Air Lock Valves	0.037			
14 - Vibrating Feeders	0.091			
56 - Vibrating Tables	0.657			
Section Subtotal =	0.915	3	2.74	2.74
SPENT BED MATERIAL COOLING				
2 - Air Lock Valves @ 15 TPH	0.027			
2 - Surge Bins @ Cooler	0.042			
4 - Air Lock Valves @ 30 TPH	0.068			
2 - Solids Cooler & Cyclones	0.312			
Section Subtotal =	0.449	2	0.90	0.90
SOLIDS HANDLING CONTROLS				
1 - Control System				
Section Subtotal =	0.149	3	0.45	0.45
OFF-SITE SPENT SOLIDS STORAGE			0.0	3.79
Total for Solids Handling			9.45	13.24

Table B5

HOT GAS AND AIR MAJOR EQUIPMENT FOR AFBC

HOT GAS CLEANUP AND AIR SUPPLY	GE Cost Estimate, 1975 M$	Number	Original GE Estimate, M$	Estimate Used in Base Plant Design, M$
4 - Bed Cyclone Units	0.392			
12 - Cyclone Air Lock Valves	0.078			
4 - Fines Injection Systems	0.260			
2 - CBC Cyclone Units	0.073			
2 - CBC Cyclone Air Lock Valves	0.026			
2 - Surge Bins @ Dust Cooler	0.040			
2 - Coolers for CBC Dust	0.260			
4 - Cooler Air Lock Valves	0.051			
4 - Electrostatic Precipitators*	2.406			
1 - Baghouse Filter System*	(2.406)			
2 - Air Preheaters	2.393			
2 - ID Fans and Motors	0.881			
2 - FD Fans and Motors	0.965			
2 - PA Fans and Motors	0.091			
Section Subtotal =	7.916	2	15.83	15.83

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*Base design substitutes a baghouse filter system for the original electrostatic precipitators and assumes identical costs for them.

Table B6

TOWER COMPONENTS FOR AFBC

Subsystems Items	GE Cost Estimate, 1975 M$	Number	Original GE Estimate, M$	Estimate Used in Base Plant Design,* M$
Heat Exchange and Pressure Parts	5.063			
Injector and Air Parts of AFB	0.781			
Control System	0.724			
Flues, Ducts, Insulation, etc.	0.893			
	7.461	4	29.84	29.84

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*No changes made in the base tower components.

Table B7

AFBC BASE PLANT CAPITAL COST BREAKDOWN*

Costs (Millions of Dollars)

Categories	Major Components	Direct Labor(1)	Indirect Field(2)	Balance-of-Plant Materials(3)	Total
Steam Generators	45.68	10.16	9.15	6.30	71.29
Turbine Generator	27.00	1.53	1.37	0.10	30.00
Process Mechanical Equipment	13.24 (9.45)	6.17	5.55	29.40	54.36 (50.57)
Electrical		12.34	11.10	12.30	35.74
Civil and Structural		10.40	9.36	13.70	33.46
Process Piping and Instrumentation		8.28	7.46	10.10	25.84
Yardwork and Miscellaneous		1.59	1.43	1.70	4.72
	85.92 (82.13)	50.47	45.42	73.60	255.41 (251.62)
BOP Labor, Materials & Indirects (Sum of 1 + 2 + 3)					169.49
A/E Home Office & Fee @ 15%					25.42
Contingency @ 20%					56.17 (55.41)
Escalation & Interest during Construction					184.68 (182.18)
Total, M $					521.68 (514.63)
Total, $/kWe					640.7 (632.0)

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*GE original cost estimate in parentheses where they differ from the base costs.

Table B8

MAJOR COMPONENT COSTS COMPARISONS IN PFBC (From Ref. B8)

Subsystem	Cost. $/kWt[a]	
	General Electric	Westinghouse
Coal and Dolomite Handling:		
Injection	9.58	1.70
Preparation and other	2.46	1.47
Subtotal	12.04	3.17
Hot-Gas Cleanup[b]:		
Granular bed filters	18.11	2.81
Cyclones and other	9.85	3.31
Subtotal	27.96	6.12
PFB Furnace Modules[c]	6.36	4.34
Gas Turbine-Generators	11.11	9.71
Stack Gas/Feedwater Heat Exchangers	1.09	3.74
Miscellaneous[d]	1.93	2.27
Total	60.50	29.36

[a] kWt is the kilowatts of thermal power associated with the HHV of the input coal. The thermal power inputs are 2307 MWt for the GE case and 1742 MWt for the Westinghouse case.

[b] Excluding hot-gas piping from major-component cost.

[c] Excluding auxiliary air compressors.

[d] Including auxiliary air compressors, solids waste handling, etc.

Table B9

SOLIDS HANDLING MAJOR EQUIPMENT FOR PFBC

Subsystem	Estimate, 1975 M$	Number	GE Original Cost Estimate, M$	Estimate Used In Base Plant Cost, M$
COAL PROCESSING & FEEDING				
1 - Dryer System @ 182-1/2 TPH	0.884	3		
1 - Crusher & 2 Screens	0.159	3		
1 - Distribution Bin @ Crusher	0.031	3		
2 - Vibrating Feeders @ 92 TPH	0.012	3		
2 - Surge Bins @ 7600 ft^3	0.060	3		
2 - Bin Activators	0.046	3		
2 - Feeders (vibrating) @ 365 TPH	0.014	3		
2 - Coal Distribution Boxes	0.010	3		
2 - Petrocarb Coal Injector System	7.198	2		2.555
Section Subtotal =	8.414		18.04	6.20
DOLOMITE PROCESSING & FEEDING				
1 - Dryer System @ 84 TPH	0.468	3		
1 - Crusher & 2 Screens	0.093	3		
1 - Distribution Bin @ Crusher	0.015	3		
2 - Vibrating Feeders @ 42 TPH	0.010	3		
2 - Surge Bins @ 1867 ft^2	0.028	3		
2 - Bin Activators	0.042	3		
2 - Feeders (vibrating)	0.013	3		
2 - Dolomite Distribution Boxes	0.008	3		
2 - Petrocarb Injection Systems	3.851	2		1.367
Section Subtotal =	4.528		9.73	3.40
SPENT BED MATERIAL				
2 - Throttle Valves for Bed Drains	0.024	2		
2 - 190 ft^3 Surge Hoppers	0.087	2		
4 - Lockhopper Valves	0.069	2		
2 - 580 ft^3 lockhoppers	0.170	2		
2 - High Temp. Feeders (vibrating) 34 TPH	0.029	2		
2 - Surge Bins @ Solids Coolers	0.040	2		
2 - Solids Coolers with Fans and Cyclones for 24 MBtu/hr	0.299	2		
4 - Air Lock Valves for Coolers @ 34 TPH	0.073	2		
Section Subtotal =	0.791		1.58	1.58
CONTROL SYSTEMS				
Solids Handling Equipment	0.149	2		
Section Subtotal =	0.149		0.30	0.30
OFF-SITE SPENT SOLIDS STORAGE	0.000		0.00	4.845
Total for PFBC Solids Handling			29.65	16.33

Table B10

FACTORS THAT AFFECT COMPARISON OF SOLIDS INJECTION WITH PETROCARB SYSTEMS (From Ref. B8)

	General Electric	Westinghouse[a]
Mode of Injecting Coal and Dolomite	Separate	Mixed
Bed Area per Injector(s)	6 ft^2/injector pair	23 ft^2/injector
Number of Injectors per Petrocarb Unit	22	4
Number of Petrocarb Units	24	16
Component Cost Per Injection Rate, \$/(lb/hr)	22	4

[a] Westinghouse has increased the number of injectors from 4 to 8. With 8 injectors/Petrocarb unit, the bed area to be served by an injector becomes 11.5 ft^2. Westinghouse estimated that the cost associated with this change is negligible.

Off-site spent solids storage in a diked, clay-lined impoundment is also included in the base design (Table B9) as specified and estimated by TVA.

Design factors for the granular bed filter system are compared in Table B11, the major cost factors being the higher face velocity selected in the Westinghouse design and the fact that GE used a multitude of small Ducon filter units (16 per boiler module) relative to the number used in the Westinghouse design (4 per boiler module)--see Table B2.

Table B12 shows the cost breakdown in the hot gas cleanup section (part of the PFB steam generator category). Both the original GE cost estimate and the base plant estimates are listed. Also shown are the high-pressure air equipment cost breakdown and the heat exchange equipment (Table B13), where no cost modifications have been performed. Also no changes in balance-of-plant costs (direct labor, indirect field costs, and materials) have been made.

The total PFBC power plant cost is broken down by categories in Table B14, with both the original GE costs and the base plant costs being listed. The TVA cost adders for cost uncertainties in developmental components were not included either for AFBC or PFBC.

In the parametric studies performed in this study, the performance values and operating conditions are modified from the base values in Table B3, as discussed in the body of this report. Equipment design changes that occur as a result of these modifications are accounted for in the cost model (see Appendix C) to the degree of accuracy required here. The assumptions applied in the equipment design changes are based largely on the finding of previous Westinghouse conceptual design studies. The base designs selected in Table B3 and the resulting cost projections are believed to be meaningful and of general significance because of the general agreement found between the existing comprehensive conceptual designs when they are placed on the same cost basis and when the technical differences are evaluated.

Table B11

DIFFERENCES IN HOT-GAS CLEANUP WITH CYCLONE SEPARATORS AND DUCON FILTERS (From Ref. 8)

	General Electric	Westinghouse
Combustion Gas from Main Bed Cell and Carbon Burnup Cell:		
Particulate loading, ppm	43,900	37,500
Particle size distribution	More small particles	More large particles
Ducon Filters:		
Particulate loading, ppm	1,790	750
Face velocity, ft/min	39	75
Bed area per volume flow, ft^2 per actual ft^3/min	0.026	0.013
Component cost, dollars per actual ft^3/min	60.5	8.9

Table B12

HOT GAS CLEANUP AND AIR MAJOR EQUIPMENT FOR PFBC

	GE Cost Estimate, 1975 M$	Number	GE Original Estimate, M$	Estimate Used in Base Plant Cost, M$
HOT GAS CLEANUP				
16 - Two-in-One Cyclones for Beds	$8.128			
4 - 145 ft^3 Collecting Hoppers	0.107			
32 - Trickle Valves & Dip Legs	0.229			
8 - Lock Hopper Seal Valves	0.137			
4 - 290 ft^3 Lock Hoppers	0.173			
4 - Fines Injection Systems	0.520			
4 - Injection System Valves	0.069			
2 - Two-in-One Cyclones for CBC	0.532			
2 - 290 ft^3 Collecting Hoppers	0.087			
4 - Trickle Valves & Dip Legs	0.029			
4 - Lock Hopper Seal Valves	0.069			
2 - 580 ft^3 Lock Hoppers	0.170			
2 - High Temp. Feeders (Vibrating)	0.031			
2 - Surge Bins for Dust Coolers	0.040			
4 - Airlock Valves for Dust Coolers	0.057			
2 - CBC Dust Coolers	0.260			
32 - Granular Bed Filters	20.894			7.339
4 - Fines Letdown & Removal Systems for Granular Beds	0.721			
Section Subtotal =	32.253	2	64.51	30.06
HIGH PRESSURE AIR				
2 - Air Dryer Precoolers	0.174	2		
4 - Air Dryers	0.350	2		
4 - Booster Compressors	0.416	2		
4 - 400 ft^3 Receivers 800#	0.051	2		
4 - Air Compressors for Granular Beds	0.193	2		
2 - 400 ft^3 Receivers 400#	0.029	2		
1 - Booster Compressor Spare	0.104	1		
1 - Air Compressor Spare	0.048	1		
Section Subtotal =	1.365		2.58	2.58

Table B13

HEAT EXCHANGE MAJOR EQUIPMENT FOR PFBC

	GE Cost Estimate, M$ per Module	GE Original Estimate, M$	Estimate used in Base Plant Cost, M$
PFB Heat Exchange and Pressure Parts	2.213		
PFB Containment Shell and Nozzles	0.566		
PFB Fuel Injection and Air Parts	0.237		
PFB Controls	0.567		
PFB Petrocarb Cooler Economizer E_x	0.089		
PFB Module	3.67	14.68	14.68
Gas Turbine Economizer E_Y	0.627	2.51	2.51

Table B14

PFBC BASE PLANT CAPITAL COST BREAKDOWNS*

Categories	Major Components	Costs, M$			
		Site Labor			
		Direct Labor (1)	Indirect Field (2)	Balance-of-Plant Materials (3)	Total
PFB Steam Generators	44.74 (79.19)	4.35	3.91	3.10	56.10 (90.55)
Turbine Generators	50.62	1.70	1.53	0.20	54.05
Process Mechanical Equipment	21.42 (34.74)	6.29	5.66	26.20	59.57 (72.89)
Electrical		9.52	8.57	11.00	29.09
Civil and Structural		9.99	8.99	11.20	30.18
Process Piping and Instrumentation		13.51	12.16	20.10	45.77
Yardwork and Miscellaneous		1.59	1.43	1.70	4.72
	116.78 (164.55)	46.95	42.25	73.50	279.48 (327.25)
		B.O.P. Labor, Materials & Indirects (sum of 1 + 2 + 3)			162.70
		A/E Home Office & Fee @ 15%			24.41
		Contingency @ 20%			60.78 (70.33)
		Escalation & Interest during Construction			199.84 (231.25)
		Total (M$)			564.5 (653.2)
		Total ($/kWe)			624.6 (722.7)

*General Electric cost estimates in parentheses where they differ from the base plant costs.

REFERENCES

B1. Archer, D. H. et al., "Evaluation of the Fluidized-Bed Combustion Process", Vol. II. Report to Office of Air Programs, EPA, by Westinghouse Research and Development Center, Pittsburgh, PA., November 1971, Contract 70-9, NTIS PB 212-916.

B2. Energy Conversion Alternatives Study (ECAS) General Electric Phase II Final Report, Volume II, NASA CR-134949, 1977, NTIS PB 269-379.

B3. Garner, D. N., et al., "A Comparison of Selected Design Aspects of Three Atmospheric Fluidized Bed Combustion Conceptual Power Plant Designs," Radcon Corp., presented at the 5th International Fluidized Bed Combustion Conference, Washington, DC, December 1977.

B4. Keairns, D. L., et al., "Fluidized-Bed Combustion Process Evaluation-Phase II- Pressurized Fluidized-Bed Coal Combustion Development," Report to EPA, Westinghouse Research Laboratories, Pittsburgh, PA, September 1975, EPA-650/2-75-027c, NTIS PB 246-116.

B5. Energy Conversion Alternatives Study (ECAS) Westinghouse Phase II Final Report, Volume III, NASA CR-134942, 1977, NTIS PB 268-558.

B6. Keairns, D. L., et al., "Evaluation of the Fluidized-Bed Combustion Process," Vol. I and II. Report to Office of Research and Development, EPA, Westinghouse Research Laboratories, Pittsburgh, PA., December 1973, EPA-650/2-73-048a and b, NTIS PB 231-162 and PB 231-163.

B7. Utility Boiler Design/Cost Comparison: Fluidized-Bed Combustion versus Flue Gas Desulfurization. Report to EPA, Tennessee Valley Authority, November 1977, EPA-600/7-77-126.

B8. Evaluation of Phase 2 Conceptual Designs and Implementation Assessment Resulting from the Energy Conversion Alternatives Study (ECAS), submitted by NASA, Lewis Research Center, to ERDA and NSF, April 1977, NTIS PB 270 017.

APPENDIX C

RELATIONSHIPS FOR FBC COST EVALUATIONS

First-order approximations have been used to scale the costs and performance of AFBC and PFBC power plants. The base plant costs (Appendix B) have been broken down to permit scaling with respect to several parameters: sorbent calcium-to-sulfur ratio, coal properties (sulfur content, ash content, heating value), sorbent properties (molecular weight, mass fraction weight reduction on calcination, heat of calcination), cost of sorbent, and cost of coal. Both equipment size/capital costs and operating costs are considered.

ATMOSPHERIC-PRESSURE FLUIDIZED-BED COMBUSTION

In order to scale the cost of an AFBC power plant with respect to the parameters listed, the rate of coal consumption, the rate of fresh sorbent feeding, the rate of spent solids, and the net plant power are estimated with respect to the base values of these quantities. For the AFBC power plant the base value of the coal-energy input rate, E, is fixed for all the cases considered at 2.26 GJ/s (7713 x 10^6 Btu/hr). Thus, the coal rate, F_c, will be

$$F_c = \frac{E}{H} \quad , \tag{1}$$

where H is the coal heating value (as received, wet coal). The fresh sorbent rate, F_s, is

$$F_s = (Ca/S) \frac{X_s}{32} F_c M \quad , \tag{2}$$

where X_s is the coal sulfur-content (weight fraction), M is the sorbent molecular weight (sorbent weight per mole of calcium) and Ca/S is the molar calcium-to-sulfur ratio. The rate of spent solids, F_d, disposal is given approximately by

$$F_d = F_c \left\{ (\mathrm{Ca/S}) \frac{X_s}{32} m \; (1 - X_{CO_2}) + X_s \frac{80}{32} + X_A \right\} , \qquad (3)$$

where X_{CO_2} is the weight fraction of the sorbent released as CO_2 on calcination, and X_A is the coal ash-content (weight fraction). For simplicity, in this expression for F_d all of the coal sulfur is assumed to be found in the spent sorbent or coal ash as calcium sulfate.

The net plant power output, P, is given approximately by

$$P = P_b - \frac{\eta(f_b - f)}{3414.4} F_c \; H - \frac{\eta \; H_R(F_s - F_{s,b})}{m \; 3414.4} - \Delta P_A \quad , \qquad (4)$$

where P_b is the base net-power, η is the gross generation efficiency for the plant, and f and f_b are the combustion efficiencies for the parametric plant and the base plant, respectively. H_R is the sorbent heat of calcination, $F_{s,b}$ is the fresh sorbent feed rate for the base plant, and ΔP_A is the change in auxiliary power between the base plant and the parametric plant. In developing equation 4, several simplifying assumptions were applied. For example, the combustor exit temperature, the steam superheat and reheat temperatures and the stack gas temperature are assumed identical with the base case values for all parametric cases. Other exothermic or endothermic reactions occurring in the combustor are assumed to generate energy at the base rates for all parametric cases. For example, the exothermic sorbent sulfation reaction that occurs in the combustor is assumed to generate energy at the base rate for 83 percent sulfur removal from a 3.9 wt % sulfur coal. Higher sulfur-content coals and/or higher sulfur removal efficiencies would actually increase the plant power slightly, but this represents a small contribution. Energy losses, above the base value, from the disposal of hot spent-solids are neglected, with a hot-air energy recovery system for coal and sorbent drying assumed to function with fixed thermal efficiency under all parametric conditions. System pressure-drops, such as the bed pressure drop, are assumed not to change significantly for the set of parameters considered.

The specifications applied in this study are as follows:

- The coal higher heating value, for high-sulfur Eastern coal of rank similar to the base Illinois No. 6 coal and containing about 13 wt % moisture, is expressed as

$$H = 13{,}922\ (1 - X_A) + 4{,}100\ X_s - 1{,}810\ \text{(Btu/lb)} \qquad (5)$$

- For a typical limestone that might be applied for AFBC, $X_{CO_2} = 0.41$, $H_R = 78{,}700$ Btu/mole, and $m = 111.0$ lb/mole calcium.

- Base values for AFBC are (from the General Electric ECAS study[C2] and Appendix B)

$F_{c,b} = 7$ [illegible] 78×10^5 lb/hr

$F_{s,b} = 1.9096 \times 10^5$ lb/hr

$F_{d,b} = 2.4923 \times 10^5$ lb/hr

$P_b = 814{,}000$ KWe

$f_b = 0.984$

$\eta = 0.439$.

- Auxiliary power consumption terms, as listed by General Electric, are presented in Table C1. The primary air fans (PA fans) and the two solids handling terms are sealed according to the total coal and sorbent rate to yield an auxiliary power term of

$$\Delta P_A = 3{,}800 \left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1 \right) + \Delta P_p\ \text{(kWe)} \quad , \qquad (6)$$

Table C1

AUXILIARY LOSS BREAKDOWN FOR AFBC (from Ref. C2)

Item	Assumptions	No. of Units	Total, MWe
Furnace			
SA Fans	24" Δ P, 0.82 eff	4	0.8
FD Fans (4)	65" Δ P, 0.82 eff	4	21.1
PA Fans (4)*	16" Δ P, 0.82 eff	4	0.5
ID Fans (4)	16" Δ P, 0.78 eff	4	5.3
ESP (8)		8	3.3
Solids handling*		4	1.4
			32.5
Turbine	0.33% of Gross kW	1	2.9
Auxiliaries			
Major Pumps			
Service	A/E estimate	2	0.9
Booster	600 psi, 5.44 x 10^6 lb, 75% x 90%	2	3.4
Condensate	185 psi, 4.86 x 10^6 lb, 70% x 90%	2	1.1
Circ. water	Proportion to cooling heat duty	3	4.7
			10.1
Water Intake	A/E estimate	2	1.0
Solids Handling*	Based on rates and lifts	1	1.9
"Hotel" Loads	A/E estimate 1% of generation	1	8.8
Cooling Tower Fans	Proportional to heat duty	24	2.8
Transformers	0.5% of gross generation	4	4.4
		TOTAL AUXILIARY POWER =	64.4

*Auxiliary power scaled for these items.

where ΔP_p represent additional auxiliary power for particulate control above the base value.

- The net plant power becomes, from equation 4,

$$P = 831{,}408 - \Delta P_p - 3{,}800\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1\right) - 0.3162(\mathrm{Ca/S})\, X_s F_c \;. \quad (7)$$

In equation 7 it is assumed that the combustion efficiency, f in equation 4, remains fixed at the base value, f_b, for all values of the parameters considered. Various models of the combustion losses were applied to express f as a function of the design and operating parameters. For example, if the combustion rate is first order with respect to the carbon content of the bed, if the CBC is increased in size in direct proportion to the carbon overflow rate, and if carbon elutriation can be neglected, then it can be shown that

$$f_b - f = \left(\frac{F_d}{F_{d,b}} - 1\right)(1 - f_b) \;, \quad (8)$$

and as the sorbent feed rate increases the combustion efficiency drops. Since carbon elutriation is believed to be the main mechanism of carbon losses, the above model is probably a poor approximation. It is also found when applying this model that the carbon losses are negligible in impact on the plant power when compared to the impact of the calcination energy loss occurring with increased calcium-to-sulfur ratio. Because the calcination energy loss is so dominant, the assumption of constant combustion efficiency is justified in these parameteric studies.

- If the sorbent particle diameter is reduced from the base value of about 1000 μm to an average of 500 μm, an additional 1320 kWe is subtracted from the base power to account for power requirements for additional grinding; the auxiliary power than becomes

$$\Delta P_A = 3278\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1\right) + 1320\left(\frac{F_s}{F_{s,b}} - 1\right) + \Delta P_p \;, \quad (9)$$

resulting in

$$P = 830{,}088 - \Delta P_p - 3278\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1\right) - 1{,}320\left(\frac{F_s}{F_{s,b}} - 1\right) - 0.3162(\mathrm{Ca/S})\, X_s F_c \quad . \qquad (10)$$

These simple and approximate relationships express the required major scaling quantities as a function of the parameters of interest.

The General Electric costs are described in a report prepared by TVA,[C4] and these have been applied with several modifications in this evaluation (see Appendix B). Seven plant categories from the ECAS cost format are used to summarize the FBC capital investment breakdown, and five of these categories are assumed to be fixed in cost over the range of parameters considered (turbine generators, electrical, civil and structural, process piping and instrumentation, and yardwork and miscellaneous), while two categories are scaled according to changes in capacity (steam generators and process mechanical equipment).

Component costs, direct labor costs, indirect field costs and materials costs are scaled according to the applicable capacity ratios $F_c/F_{c,b}$, $(F_c + F_s)/(F_{c,b} + F_{s,b})$ $F_{s,b}$ and $F_d/F_{d,b}$, with a 0.85 power factor for solids handling items, with a 0.6 power factor for cyclones and auxiliary components, and a 0.68 power factor for fans and motors. These factors were selected from previous plant cost studies performed by Westinghouse.

The process mechanical equipment category includes the scaled items coal drying of crushing, limestone drying and crushing, coal and limestone blending and feeding, spent bed material cooling, and spent bed material off-site storage. Solids handling controls are assumed to be constant in cost. The investment to haul spent solids off site and store them in a clay-lined, diked impoundment is taken from TVA projections.

The steam generator category includes hot gas cleanup items, air supply, and the combustor and steam generator items. The hot gas cleanup items that are assumed to be unchanged in cost by the study parameters are the cyclone units for the primary combustion beds, cyclone air lock valves, fines injection system, carbon burnup cyclone air lock valves, cooler air lock valves, and the baghouse filter system. The carbon burnup cell cyclone units, the surge bins and dust cooler, and the cooler for the carbon burnup cell dust are scaled according to capacity. The additional cost of increased particulate control capability above the base design is considered separately (Appendix D).

The PA (primary air) fans and motors, the air preheaters, the ID fans and motors, and the FD fans and motors are all scaled according to increased capacity.

The combustor steam generator is assumed to be constant in cost for the parameters of interest in this study, based on previous comments on cost sensitivity.

Tables C2 through C5 list the major equipment and balance-of-plant items, indicating which items are scaled and with what power factor they are scaled.

This method results in a parametric capital investment, I, given by

$$I(\$ \times 10^6) = 453.4 + 25.0\left(\frac{F_c}{F_{c,b}}\right)^{0.85} + 10.7\left(\frac{F_s}{F_{s,b}}\right)^{0.85} + 8.5\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}}\right)^{0.85} + 25.1\left(\frac{F_d}{F_{d,b}}\right)^{0.85} + 1.5\left(\frac{F_d}{F_{d,b}}\right)^{0.6} + 1.7\left(\frac{F_d}{F_{d,b}}\right)^{0.68} + 0.37\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}}\right)^{0.68} \quad . \qquad (11)$$

The energy cost, E_c, in mills/kWh is related to the investment, I, the net plant power, P, the coal cost, p($/MBtu) and the sorbent cost, S($/ton), by

$$E_c = \frac{3.5037 \times 10^4 \; I + p \; F_c H \times 10^{-3} + 0.5 \; S \; F_s}{P} \quad . \qquad (12)$$

The above equations for I and E_c were used in calculating the AFBC costs in Figure 4 through 6, 8, and 9 in the main text.

Table C2

SCALING OF SOLIDS HANDLING MAJOR EQUIPMENT FOR AFBC

Subsystems	Cost Estimate, 1975 M$	Number	Cost Scaling Factor Used	Estimate Used in Base Plant Cost, M$
COAL DRYING & CRUSHING				
1 - Dryer System-189 TPH	0.910			
1 - Coal Crusher & 2 Screens	0.159			
1 - Distribution Box	0.031			
2 - Vibrating Feeders @ 94-1/2 TPH	0.012			
2 - Surge Bins @ 7600 ft^3	0.060			
2 - Bin Activators	0.021			
2 - Weigh Belt Feeders @ 94-1/2 TPH	0.053			
Section Subtotal =	1.246	3	$(F_c/F_{c,b})^{0.85}$	3.74
LIMESTONE DRYING & CRUSHING				
1 - Dryer System @ 43.6 TPH	0.359			
1 - Crusher & 2 Screens	0.083			
1 - Distribution Box	0.011			
2 - Vibrating Feeders @ 24.3 TPH	0.008			
2 - Surge Bins	0.022			
4 - Bin Activators	0.026			
2 - Air Lock Valves	0.006			
2 - Vibrating Feeders	0.008			
2 - Weigh Belt Feeders @ 24.3 TPH	0.018			
Section Subtotal =	0.541	3	$(F_s/F_{s,b})^{0.85}$	1.62
COAL & LIMESTONE BLENDING & FEEDING				
2 - Blenders @ 115 TPH	0.130			
14 - Air Lock Valves	0.037			
14 - Vibrating Feeders	0.091			
56 - Vibrating Tables	0.657			
Section Subtotal =	0.915	3	$\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}}\right)^{0.85}$	2.74

Table C2 (Cont'd)

Subsystems	Cost Estimate, 1975 M$	Number	Cost Scaling Factor Used	Estimate Used in Base Plant Cost, M$
SPENT BED MATERIAL COOLING				
2 - Air Lock Valves @ 15 TPH	0.027			
2 - Surge Bins @ Cooler	0.042			
4 - Air Lock Valves @ 30 TPH	0.068			
2 - Solids Cooler & Cyclones	0.312			
Section Subtotal =	0.449	2	$(F_d/F_{d,b})^{0.85}$	0.90
SOLIDS HANDLING CONTROLS				
1 - Control System				
Section Subtotal =	0.149	3	None	0.45
OFF-SITE SPENT SOLIDS STORAGE				
Total for Solids Handling	0.0		$(F_d/F_{d,b})^{0.85}$	3.79
				13.24

Table C3

SCALING OF HOT GAS AND AIR MAJOR EQUIPMENT FOR AFBC

Subsystems	Cost Estimate, 1975 M$	Number	Cost Scaling Factor Used	Estimate Used in Base Plant Design, M$
HOT GAS CLEANUP AND AIR SUPPLY				
4 - Bed Cyclone Units	0.392			
12 - Cyclone Air Lock Valves	0.078			
4 - Fines Injection Systems	0.260			
2 - CBC Cyclone Units	0.073		$(F_d/F_{d,b})^{0.6}$	
2 - CBC Cyclone Air Lock Valves	0.026			
2 - Surge Bins @ Dust Cooler	0.040		$(F_d/F_{d,b})^{0.6}$	
2 - Coolers for CBC Dust	0.260			
4 - Cooler Air Lock Valves	0.051			
1 - Baghouse Filter System	2.406			
2 - Air Preheaters	2.393		$[0.9 + 0.1 (F_d/F_{d,b})]^{0.68}$	
2 - ID Fans and Motors	0.881			
2 - FD Fans and Motors	0.965			
2 - PA Fans and Motors	0.091		$[(F_c + F_s)/(F_{c,b} + F_{s,b})]^{0.68}$	
Section Subtotal =	7.916	2		15.83

Table C4

TOWER COMPONENTS FOR AFBC

Subsystems Items	M$	Number	Cost Scaling Factor Used	Estimate Used in Base Plant Design, M$
Heat Exchange and Pressure Parts	5.063			
Injector and Air Parts of AFB	0.781			
Control System	0.724			
Flues, Ducts, Insulation, etc.	0.893			
	7.461	4	None	29.84

PRESSURIZED FLUIDIZED-BED COMBUSTION

The development for PFBC is similar to that for AFBC. For PFBC the coal energy input rate is fixed as

$$E = F_c H = 7874 \times 10^6 \text{ (Btu/hr)} \tag{13}$$

from the General Electric ECAS conceptual design. Relationships 2, 3, and 4 are applied for PFBC with a dolomite sorbent having characteristics given by $X_{CO_2} = 0.474$, $H_R = 100{,}800$ Btu/mole and m = 188.0 lb/mole calcium. The base values for PFBC are assumed to be

$$F_{c,b} = 7.2005 \times 10^5 \text{ lb/hr}$$

$$F_{s,b} = 3.2996 \times 10^5 \text{ lb/hr}$$

$$F_{d,b} = 3.1289 \times 10^5 \text{ lb/hr}$$

$$P_b = 904{,}000$$

$$\eta = 0.465$$

$$f_b = 0.985 \quad .$$

In these relations the combustor temperature has been raised from the General Electric base value of 899°C to 954°C, while all other factors are assumed to remain constant. Previous design studies have shown this to have little impact on the power plant thermal efficiency, but the higher temperature yields conditions under which the dolomite sorbent will be fully calcined in the combustor. At 899°C and the base excess air level calcination would not be expected, based on thermodynamics.

The auxiliary power term is given (based on Table C6) by

$$\Delta P_A = 11{,}450 \left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1 \right) + \Delta P_p \quad , \tag{14}$$

Table C5

SCALING OF BALANCE-OF-PLANT COSTS FOR AFBC

		Direct Manual Field Labor MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
1.0	AFB STEAM GENERATORS (3)			
1.1	Steam Generator Erection			
-	Erect only (supply by others): includes heat transfer surface and pressure parts; buckstays, braces and hangers; attached flue and duct; fluid-bed components; control equipment and valves; miscellaneous piping and small valves; insulation for above	343		None
-	Supply and erect: includes support steel for above; access steel; miscellaneous materials and labor operations	158	3,020	None
1.2	Steam Generator Auxiliaries			
-	Erect only (supply by others): includes PA fans; FD fans; ID fans; air preheaters; coal and limestone feed tables; pipes and air lock valves	70	100	None
-	Supply and erect: includes external flue and duct; support steel for ductwork; insulation for external ductwork	198	2,570	None
1.3	Stack Gas Cleanup:			
-	Erect only (supply by others): includes cyclones and electrostatic precipitators	82		None
-	Supply and erect: includes support steel for cyclones and precipitators	14	610	None
		865	6,300	
2.0	TURBINE GENERATOR (3)			
-	Install only (supply by others): includes 883 MWe steam turbine; generator; exciter; auxiliary equipment; integral steam and auxiliary piping; insulation; miscellaneous labor operations	130	100	None
3.0	PROCESS MECHANICAL EQUIPMENT (3)			
3.1	Boiler Feedwater Pumps			
-	Supply and install: includes turbine-driven main feedwater pumps and drivers (3 @ $940,000); feedwater booster pumps and motors (2 @ $125,000)	10	3,200	None
3.2	Main Circ. Water Pumps (3)			
-	Supply and install: includes main circ. water pumps and motors (3 @ $250,000)	3	790	None

Table C5 (Cont'd)

		Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
3.3	Other Pumps (3)			
-	Supply and install: includes condensate pumps and motors (2 @ $100,000); and other pumps and drivers not listed elsewhere	5	670	None
3.4	Main Condenser (3)			
-	Supply and install: includes shells; tubes; air ejectors	27	2,520	None
3.5	Heaters, Exchangers, Tanks and Vessels (3)			
-	Supply and install: includes l.p. feedwater heaters (4); i.p. feedwater heater; h.p. feedwater heater; deaerating heater and storage tank; miscellaneous heaters and exchangers; tanks and vessels	7	2,580	None
3.6	Stacks and Accessories (3)			
-	Supply and erect: includes concrete stacks and liners; lights and marker painting; hoists and platforms; stack foundations	110	1,700	None
3.7	Turbine Hall Crane (1)			
-	Supply and erect: includes crane and accessories	3	420	None
3.8	Coal Handling (2)			
-	Erect only (supply by others): includes coal dryers (3); support and access steel for dryers; coal grinders (3); screens at grinders	22	10	$(F_c/F_{c,b})^{0.85}$
-	Supply and erect: includes railcar dumping equipment; dust collectors; primary crushing equipment; belt scale; sampling station; magnetic cleaners, mobile equipment; conveyors to pile; reclaiming feeders; conveyors to cascade; coal cascade; conveyors and bucket elevators to dryers and grinders; recirculating conveyors at grinders; conveyors to blenders	73	5,700	$(F_c/F_{c,b})^{0.85}$
3.9	Limestone Handling (2)			
-	Erect only (supply by others): includes limestone dryers (3); support and access steel for dryers; limestone grinders (3); screens at grinders; limestone surge bins at AFB modules	11	10	$(F_s/F_{s,b})^{0.85}$

Table C5 (Cont'd)

		Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
-	Supply and erect: includes magnetic cleaners; conveyor to limestone pile; reclaiming feeders; conveyors to cascade; limestone cascade; conveyors and bucket elevators to dryers and grinders; recirculating conveyors at grinders; conveyors to blenders; pneumatic transport feeders, hoppers, blowers, and piping to AFB modules	45	1,960	$(F_s/F_{s,b})^{0.85}$
3.10	Coal and Limestone Blend Handling (2)			
-	Erect only (supply by others): includes blenders (3); surge bins; bin unloaders and feeders	4	10	$\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}}\right)^{0.85}$
-	Supply and erect: includes conveyors and bucket elevators to AFB modules	13	920	$\left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}}\right)^{0.85}$
3.11	Spent Solids Handling (2)			
-	Erect only (support by others): includes high-temperature vibrating feeders and conveyor to solids cooler; solids cooler	6	10	$(F_d/F_{d,b})^{0.85}$
-	Supply and erect: includes fly-ash handling system for precipitators and air preheater; solids cooler accessories and bucket elevator; ash conveyors; ash storage silos (6) with feeders, unloaders and foundations; railcar loading equipment	93	4,480	$(F_d/F_{d,b})^{0.85}$
3.12	Cooling Towers (3)			
-	Supply and erect: includes mechanical draft towers with fans and motors	63	2,680	None
3.13	Other Mechanical Equipment (3)			
-	Supply and erect: includes water treatment and chemical injection; air compressors and auxiliaries; fuel oil ignition and warm-up; screenwell; miscellaneous plant equipment; equipment insulation	30	1,720	None
		525	29,400	
4.0	ELECTRICAL (5)			None
4.1	Main Transformers	4	2,030	
4.2	Other Transformers and Main Bus	15	1,160	
-	includes start-up transformer; station service transformers; generator main bus			
4.3	Switchgear and Control Centers	37	3,020	
-	includes switchgear and load centers; motor control centers; local control stations; distribution panels, relay and meter boards			

Table C5 (Cont'd)

		Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
4.4	Other Electrical Equipment	428	2,400	
-	includes communications, grounding; cathodic and freeze protection; lighting; preoperational testing			
4.5	Auxiliary Diesel Generator	2	110	
-	includes diesel generator, batteries and associated dc equipment			
4.6	Conduit, Cable Trays, Wire and Cable	564	3,580	
		1,050	12,300	
5.0	CIVIL AND STRUCTURAL			None
5.1	Concrete Substructures and Foundations (1)	350	2,890	
-	includes turbine building substructure; AFB base mats; coal, limestone and ash handling foundations, pits and tunnels; miscellaneous equipment foundations; auxiliary buildings substructures; miscellaneous concrete			
5.2	Superstructures (1)	230	6,540	
-	includes turbine building; auxiliary yard buildings			
5.3	Earthwork (1)	135	300	
-	includes building excavations; coal, limestone and ash handling excavations; circ. water system excavations; AFB foundation excavations; miscellaneous foundation excavations; dewatering and piling			
5.4	Cooling Tower Basin and Circ. Water System (3)	115	1,800	
-	includes circ. water pumps pads, riser and concrete envelope for pipe; cooling tower basin; circ. water pipe; cooling tower miscellaneous steel and fire protection			
5.5	AFB Boiler Enclosures (1)	55	2,170	
-	includes structural steel; noninsulated walls and roofing; building services; elevators			
		885	13,700	
6.0	PROCESS PIPING AND INSTRUMENTATION			None
6.1	Steam and Feedwater Piping (3)	55	2,680	
-	includes main steam; extraction steam; hot reheat; cold reheat; feedwater and condensate large piping, valves and fittings			
6.2	Other Large Piping (3)	165	2,670	
-	includes auxiliary steam; process water; auxiliary systems			

Table C5 (Cont'd)

		Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
6.3	Small Piping (3)	85	760	
-	includes all piping, valves and fittings of 2-in. diameter and less			
6.4	Hangers and Misc. Labor Operations (3)	245	920	
-	includes all hangers and supports; material handling; scaffolding; misc. labor operations			
6.5	Pipe Insulation (3)	40	430	
6.6	Instrumentation and Controls (5)	115	2,640	
		705	10,100	
7.0	YARDWORK AND MISCELLANEOUS (1)			None
7.1	Site Preparation and Improvements	38	10	
-	includes soil testing; clearing and grubbing; rough grading; finish grading; landscaping			
7.2	Site Utilities	5	50	
-	includes storm and sanitary sewers; nonprocess service water			
7.3	Roads and Railroads	27	750	
-	includes railroad spur; roads, walks, and parking areas			
7.4	Yard Fire Protection, Fences, and Gates	52	600	
7.5	Water Treatment Ponds	12	10	
-	includes earthwork; compacted-clay lining; off-site pipeline			
7.6	Lab, Machine Shop, and Office Equipment	1	280	
		135	1,700	

Table C6

AUXILIARY LOSS BREAKDOWN FOR PFBC

Item	Assumptions	No. of Units	Total MWe
Furnace			
Fans for dryers (coal dolomite, solids)		2	4.18
Coal crusher		2	0.74
Dolomite crusher		2	0.33
Injector compressor		8	4.03
Filter		8	1.11
Turbines	0.33% of steam turbine, kW	1	2.37
	1% of gas turbine, kW	4	2.16
Pumps			
Condensate	Δ P = 185 psia	2	9.95
Circ. water	Proportion to cooling duty	3	4.40
Service water	A/E estimate	2	0.89
Intake water	A/E estimate	2	0.94
Solids Handling	Based on rates and lifts	1	2.17
"Hotel" Load	0.88% of generated kW	1	8.34
Cooling Tower Fans	Vendors value	22	2.53
Transformer Loss	0.5% of generated kW	2	4.72
		TOTAL AUXILIARY POWER =	39.86

*Auxiliary power scaled for these items.

and the plant power is given by

$$P = 928{,}094 - \Delta P_p - 11{,}450 \left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1 \right) - 0.429(\text{Ca/S})\, X_s F_c \qquad (15)$$

for the base sorbent particle size of about 2000 μm. In relation 15 it has been assumed that the combustion efficiency remains fixed at its base value for all parametric cases, while significant increases in the combustion efficiency above the base value of 98.5% are expected.

For the case of finer sorbent particles of 500 μm average diameter, modifications to the power requirements for grinding (1320 kWe) result in

$$P = 926{,}774 - \Delta P_p - 10{,}130 \left(\frac{F_c + F_s}{F_{c,b} + F_{s,b}} - 1 \right) - 1320 \left(\frac{F_s}{F_{s,b}} - 1 \right) - 0.429(\text{Ca/S})\, X_s F_c \ . \qquad (16)$$

The plant costs have been scaled, as for the AFBC case, using the General Electric cost breakdown with several modifications as detailed in Appendix B.

The scaling of cost items are summarized in Tables C7 through C9, major equipment, and C10, balance-of-plant equipment. The factors are analogous to the AFBC case.

If the modified General Electric ECAS costs are combined from these tables the plant investment, I ($ x 10^6), is given by

$$I = 498.4 + 26.4 \left(\frac{F_c}{F_{c,b}} \right)^{0.85} + 12.25 \left(\frac{F_s}{F_{s,b}} \right)^{0.85} + 24.3 \left(\frac{F_d}{F_{d,b}} \right)^{0.85} + 3.17 \left(\frac{F_d}{F_{d,b}} \right)^{0.6} \ . \qquad (17)$$

Table C7

SCALING OF SOLIDS HANDLING MAJOR EQUIPMENT FOR PFBC

Subsystems	Cost Estimate 1975 M$	Number	Cost Scaling Factor Used	Estimate Used in Base Plant Cost M$
COAL PROCESSING & FEEDING				
1 - Dryer System @ 182-1/2 TPH	0.884	3		
1 - Crusher & 2 Screens	0.159	3		
1 - Distribution Bin @ Crusher	0.031	3		
2 - Vibrating Feeders @ 92 TPH	0.012	3		
2 - Surge Bins @ 7600 ft^3	0.060	3		
2 - Bin Activators	0.046	3		
2 - Feeders (vibrating) @ 365 TPH	0.014	3		
2 - Coal Distribution Boxes	0.010	3		
2 - Petrocarb Coal Injector System	2.555	2		
Section Subtotal =	3.771		$(F_c/F_{c,b})^{0.85}$	6.20
DOLOMITE PROCESSING & FEEDING				
1 - Dryer System @ 84 TPH	0.468	3		
1 - Crusher & 2 Screens	0.093	3		
1 - Distribution Bin @ Crusher	0.015	3		
2 - Vibrating Feeders @ 42 TPH	0.010	3		
2 - Surge Bins @ 1867 ft^2	0.028	3		
2 - Bin Activators	0.042	3		
2 - Feeders (vibrating)	0.013	3		
2 - Dolomite Distribution Boxes	0.008	3		
2 - Petrocarb Injection Systems	1.367	2		
Section Subtotal =	2.044		$(F_s/F_{s,b})^{0.85}$	3.40
SPENT BED MATERIAL				
2 - Throttle Valves for Bed Drains	0.024	2		
2 - 190 ft^3 Surge Hoppers	0.087	2		
4 - Lock Hopper Valves	0.069	2		
2 - 580 ft^3 Lockhoppers	0.170	2		
2 - High-Temp. Feeders (vibrating) 34 TPH	0.029	2		
2 - Surge Bins @ Solids Coolers	0.040	2		
2 - Solids Coolers with Fans & Cyclones for 24 MBtu/hr	0.299	2		
4 - Air Lock Valves for Coolers @ 34 TPH	0.073	2		
Section Subtotal =	0.791		$(F_d/F_{d,b})^{0.85}$	1.58
CONTROL SYSTEMS				
Solids Handling Equipment	0.149	2		
Section Subtotal =	0.149		None	0.30
OFF-SITE SPENT SOLIDS STORAGE				
Total for PFBC Solids Handling	4.845		$(F_d/F_{d,b})^{0.85}$	4.845
				16.33

Table C8

SCALING OF HOT GAS CLEANUP AND AIR MAJOR EQUIPMENT FOR PFBC

	Cost Estimate 1975 M$	Number	Cost Scaling Factor Used	[illegible] in Base Plant Cost M$
HOT GAS CLEANUP				
16 - Two-in-One Cyclones for Beds	8.128			
4 - 145 ft^3 Collecting Hoppers	0.107			
32 - Trickle Valves & Dip Legs	0.229			
8 - Lockhopper Seal Valves	0.137			
4 - 290 ft^3 Lockhoppers	0.173			
4 - Fines Injection Systems	0.520			
4 - Injection System Valves	0.069			
2 - Two-in-One Cyclones for CBC	0.532		$(F_d/F_{d,b})^{0.6}$	
2 - 290 ft^3 Collecting Hoppers	0.087			
4 - Trickle Valves & Dip Legs	0.029			
4 - Lockhopper Seal Valves	0.069			
2 - 580 ft^3 Lockhoppers	0.170			
2 - High Temp. Feeders (vibrating)	0.031			
2 - Surge Bins for Dust Coolers	0.040			
4 - Airlock Valves for Dust Coolers	0.057			
2 - CBC Dust Coolers	0.260		$(F_d/F_{d,b})^{0.6}$	
32 - Granular Bed Filters	7.339			
4 - Fines Letdown & Removal Systems for Granular Beds	0.721			
Section Subtotal =	15.03	$\overline{2}$		30.06
HIGH PRESSURE AIR			None	
2 - Air Dryer Precoolers	0.174	2		
4 - Air Dryers	0.350	2		
4 - Booster Compressors	0.416	2		
4 - 400 ft^3 Receivers 800#	0.051	2		
4 - Air Compressors for Granular Beds	0.193	2		
2 - 400 ft^3 Receivers 400#	0.029	2		
1 - Booster Compressor Spare	0.104	1		
1 - Air Compressor Spare	0.048	1		
Section Subtotal =	1.365			2.58

Table C9

SCALING OF HEAT EXCHANGE MAJOR EQUIPMENT FOR PFBC

	Cost Estimate M$ per Module	Number	Cost Scaling Factor Used	Estimate Used in Base Plant Cost, M$
PFB Heat Exchange and Pressure Parts	2.213			
PFB Containment Shell and Nozzles	0.566			
PFB Fuel Injection and Air Parts	0.237		None	
PFB Controls	0.567			
PFB Petrocarb Cooler Economizer	0.089			
PFB Module	3.67	4		14.68
Gas Turbine Economizer	0.627	4		2.51

Table C10

SCALING OF BALANCE-OF-PLANT COSTS FOR PFBC

	Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
1.0 PFB STEAM GENERATORS (3)			
1.1 Steam Generator Erection			
- Erect only (supply by others): includes PFB tower skirt; PFB towers; piping connections at tower; insulation	22		None
- Supply and erect: includes tower access steel; miscellaneous materials and labor operations	6	70	None
1.2 Steam Generator Auxiliaries			
- Erect only (supply by others): includes coal and dolomite Petrocarb injection systems with injection air compressors and auxiliaries	22	100	None
- Supply and erect: includes support steel for Petrocarb systems; coal and dolomite piping from Petrocarb systems to PFB towers	221	890	None
1.3 Hot Gas Cleanup			
- Erect only (supply by others): includes cyclones; hoppers and surge bins; valves; feeders and injection systems; dust coolers; granular bed filters; granular bed blowback air compressors and auxiliaries	38	100	None
- Supply and erect: includes support steel for hot gas cleanup equipment; access steel	61	1940	None
	370	3100	
2.0 TURBINE GENERATORS (3)			
2.1 Steam Turbine Generator			
- Erect only (supply by others): includes 732 MWe steam turbine: generator; exciter; auxiliary equipment; integral steam and auxiliary piping; insulation; miscellaneous labor operations	105	100	None

Table C-10 (Cont'd)

	Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
2.2 Gas Turbine/Compressor/Generators			
- Erect only (supply by others): includes gas turbine compressors with 50 MWe generators (4)	40	100	None
	145	200	
3.0 PROCESS MECHANICAL EQUIPMENT			
3.1 Boiler Feedwater Pumps (3)			
- Supply and install: includes turbine-driven main feedwater pumps and drivers (3 @ $770,000)	8	2430	None
3.2 Main Circ. Water Pumps (3)			
- Supply and install: includes main circ. water pumps and motors (3 @ $265,000)	3	830	None
3.3 Other Pumps (3)			
- Supply and install: includes condensate pumps and motors - (2 @ $100,000); and other pumps and drivers not listed elsewhere	4	590	None
3.4 Main Condenser (3)	27	2370	None
- Supply and install: includes shells; tubes; air ejectors			
3.5 Heaters, Exchangers, Tanks, and Vessels (3)			
- Erect only (supply by others); includes heat recovery economizers; insulation; support steel (supply and erect)	20	180	
- Supply and erect: includes low-pressure feedwater heaters (2); deaerating heater and storage tank; miscellaneous heater and exchangers; tanks and vessels	7	1450	
3.6 Stacks and Accessories (3)	23	270	
- Supply and erect: includes concrete stacks and liners; lights and marker painting; hoists and platforms; foundations			

Table C-10 (Cont'd)

		Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Scaling Factor Used
3.7	Draft Ducts (3)	23	270	
-	Supply and erect: includes ducts from heat recovery economizers to stacks			
3.8	Turbine-Hall Crane (1)	3	380	
-	Supply and erect: includes crane and accessories			
3.9	Coal Handling (2)			
-	Erect only (supply by others): includes coal dryers (3); support and access steel for dryers; coal grinders (3); screens at grinders	22	10	$(Fc/Fc,b)^{0.85}$
-	Supply and erect: includes railcar dumping equipment; dust collectors; primary crushing equipment; belt scale; sampling station; magnetic cleaners; mobile equipment; conveyors to pile; reclaiming feeders; conveyors to cascade; coal cascade; conveyors and bucket elevators to dryers and grinders; recirculating conveyors at grinders; conveyors to Petrocarb; bucket elevators at Petrocarb	80	6150	$(Fc/Fc,b)^{0.85}$
3.10	Dolomite Handling (2)			
-	Erect only (supply by others): includes dolomite dryers (3); support and access steel for dryers; dolomite grinders (3); screens at grinders	13	10	$(Fs/Fs,b)^{0.85}$
-	Supply and erect: includes magnetic cleaners; conveyor to dolomite pile; reclaiming feeders; conveyors to cascade; dolomite cascade; conveyors and bucket elecators to dryers; conveyors and bucket elevators to grinders; conveyors and bucket elevators to Petrocarb	43	2240	$(Fs/Fs,b)^{0.85}$
3.11	Spent Solids Handling (2)			
-	Erect only (supply by others); includes spent solids valves; hoppers; feeders; fans; cyclones; conveyor to coolers; ash coolers (2)	11	30	$(Fd/Fd,b)^{0.85}$

Table C-10 (Cont'd)

	Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Sealing Factor Used
– Supply and erect: includes ash cooler accessories and bucket elevators; ash conveyors; ash storage silos (6) with feeders, unloaders and foundations; railcar loading equipment	85	3540	$(Fd/Fd,b)^{0.85}$
3.12 Cooling Towers (3)			
– Supply and erect: includes mechanical draft towers with fans and motors	57	2450	None
3.13 Other Mechanical Equipment (3)			
– Supply and install: includes water treatment and chemical injection; air compressors and auxiliaries; fuel oil ignition and warm-up; screenwell; miscellaneous plant equipment; equipment insulation	34	1820	None
	535	26,200	
4.0 ELECTRICAL (5)			None
4.1 Main Transformers	11	2960	
– includes main power transformers and transformers at gas turbine generators			
4.2 Other Transformers and Main Bus	18	1240	
– includes start-up transformer; station service transformers; generator main bus			
4.3 Switchgear and Control Centers	25	2050	
– includes switchgear and load centers; motor control centers; local control stations; distribution panels, relay and meter boards			
4.4 Other Electrical Equipment	373	2140	
– includes communications; grounding; cathodic and freeze protection; lighting; preoperational testing			
4.5 Auxiliary Diesel Generator	2	110	
– includes diesel generator, batteries and associated dc equipment			

Table C-10 (Cont'd)

	Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Sealing Factor Used
4.6 Conduit, Cable Trays, Wire and Cable	381	2500	
	810	11,000	
5.0 CIVIL AND STRUCTURAL			None
5.1 Concrete Substructures and Foundations (1)	395	3260	
- includes turbine building substructure; PFB base mats; coal, dolomite, and ash handling foundations, pits and tunnels; miscellaneous equipment foundations; auxiliary buildings substructures; miscellaneous concrete			
5.2 Superstructures (1)	220	6130	
- includes turbine building; auxiliary yard buildings			
5.3 Earthwork (1)	135	300	
- includes building excavations; coal, dolomite,and ash handling excavations; circ. water system excavations; PFB foundation excavations; miscellaneous foundation excavations; dewatering and piling			
5.4 Cooling Tower Basin and Circ. Water System (3)	100	1510	
- includes circ. water pump pads, riser and concrete envelope for pipe; cooling tower basin; circ. water pipe; cooling tower miscellaneous steel and fire protection			
	850	11,200	
6.0 PROCESS PIPING AND INSTRUMENTATION			None
6.1 Steam and Feedwater Piping (3)	50	2400	
- includes main steam; extraction steam; hot reheat; cold reheat; feedwater and condensate large piping, valves and fittings			

Table C-10 (Cont'd)

	Direct Manual Field Labor, MH 1000	Balance of Plant Material, $ 1000	Cost Sealing Factor Used
6.2 Hot Gas Large Pipe (3)	120	5700	
- includes PFB compressed air feed, PFB hot gas discharge, cyclone and granular bed filter piping; gas turbine inlet piping			
6.3 Other Large Piping (3)	195	3350	
- includes auxiliary steam; process water; auxiliary systems; spent solids piping			
6.4 Small Piping (3)	125	1100	
- includes all piping, valves and fittings of 2-in diameter and less			
6.5 Hangers and Misc. Labor Operations (3)	390	1900	
- includes all hangers and supports; material handling; scaffolding; misc. labor operations			
6.6 Pipe Insulation (3)	40	450	
6.7 Instrumentation and Controls (5)	230	5200	
	1,150	20,100	
7.0 YARDWORK AND MISCELLANEOUS (1)			None
7.1 Site Preparation and Improvements	38	10	
- includes soil testing; clearing and grubbing; rough grading; finish grading; landscaping			
7.2 Site Utilities	5	50	
- includes storm and sanitary sewers; non-process service water			
7.3 Roads and Railroads	27	750	
- includes railroad spurs; roads, walks and parking areas			
7.4 Yard Fire Protection, Fences and Gates	52	600	
7.5 Water Treatment Ponds	12	10	
- includes earthwork; compacted-clay lining off-site pipeline			
7.6 Lab, Machine Shop and Office Equipment	1	280	
	135	1,700	

The cost of electricity for PFBC is represented by the same general relationship used for AFBC, equation 12.

While these relationships for AFBC and PFBC are based on oversimplifying assumptions, they provide a means of understanding the major impact of the critical design and operating parameters.

REFERENCES

C1. Energy Conversion Alternatives Study (ECAS) Westinghouse Phase II Final Report, Volume III, NASA CR-134942, 1977, NTIS PB 268-558.

C2. Energy Conversion Alternatives Study (ECAS) General Electric Phase II Final Report, Volume II, NASA CR-134949, 1977, NTIS PB 269-379.

C3. Evaluation of Phase 2 Conceptual Designs and Implementation Assessment Resulting from the Energy Conversion Alternatives Study (ECAS). Report to ERDA and NRF, NASA, Lewis Research Center, April 1977, NTIS PB 270-017.

C4. Utility Boiler Design/Cost Comparison: Fluidized-Bed Combustion versus Flue Gas Desulfurization. Report to EPA, TVA, November 1977, EPA-600/7-77-126.

APPENDIX D

PARTICULATE CONTROL PROJECTIONS

In this work we have studies the effect of lower SO_2 standards on particulate effluent. By increasing the calcium-to-sulfur ratio used in fluidized-bed combustion processes, the SO_2 emissions can be reduced. A larger amount of particulate matter is carried over from the bed, however, since there is a larger amount of sorbent to be lost by attrition and elutriation per unit mass of coal burned.

A consistent methodology has been formulated and carried out for analyzing the effect of the increased carry-over on FBC flue gas particulate emissions, for both the pressurized and atmospheric cases. A discussion of the method and its assumptions follows. The ability to control these emissions to the requirement of 12.9 ng/J (0.03 lb/MBtu) is assessed.

The validity of any particulate removal study depends on the accuracy with which particle loadings and size distributions at the source can be estimated. In most previous work it has been assumed that an "average" particle represents the dust being elutriated from the bed. The properties of the three main sources of flue gas particulates in FBC (char, ash, and sorbent), however, are quite different, with the result that calculations based on such "average" particles can only be rough estimates. This work estimates loadings and size distributions for each of the three components and accounts for the differences in physical properties (density and shape) in the particulate removal calculations.

PROJECTION OF PARTICULATE EMISSIONS

Solids Feed to Fluidized Bed

The solids fed to the process consist of coal and dolomite. The coal size distribution for both the pressurized and atmospheric cases are shown as the bottom curve in Figures D1 and D2. These size distributions are typical 6350 μm and 12,700 μm distributions, respectively, from a hammer mill. The sorbent size distributions were obtained by screen analysis of an available sample of single-screened limestone from Greer. The sorbent for the pressurized case is a 6350 μm stone and for the atmospheric case it is a 3180 μm material. The very fine material from the screen analysis was analyzed with a Coulter Counter in an attempt to obtain a better estimate of the fines size distribution. These results are shown in Figures D3 and D4.

Note that the equivalent spherical diameter (d_s) has been used in these plots and that this can differ significantly from d_{avg} obtained by the screen analysis. The average particle diameter and the equivalent spherical diameter are related by the following equation:

$$d_{avg} = d_s \, (\eta\phi) \, ,$$

where η is the specific surface of the particle, and ϕ is the sphericity of the particle. For the limestone it was possible to obtain η as a function of d_{avg} by measuring the pressure drop through a bed of particles of size d_{avg}, and of known bed voidage. Ergun's equation was then used to obtain the product of ϕd_s:

$$\left(\frac{\Delta P}{L}\right) g_c = 150 \left(\frac{(1-\varepsilon)^2}{\varepsilon^3}\right)\left(\frac{\mu u_o}{(\phi d_s)^2}\right) + 1.75 \left(\frac{1-\varepsilon}{\varepsilon^3}\right)\left(\frac{\rho_g u_o^2}{(\phi d_s)}\right) .$$

The pressure drop data are shown in Figure D5. Having obtained $\eta(d_{avg})$, and assuming ϕ to be similar to sharp sand ($\phi = 0.6$), the size distribution obtained as d_{avg} with the screens could be converted and plotted as the spherical equivalent diameter (d_s). Although this analysis was carried out as carefully as possible, it should be noted that these data apply strictly only to the stone studied here and are not generally applicable to all FBC sorbents.

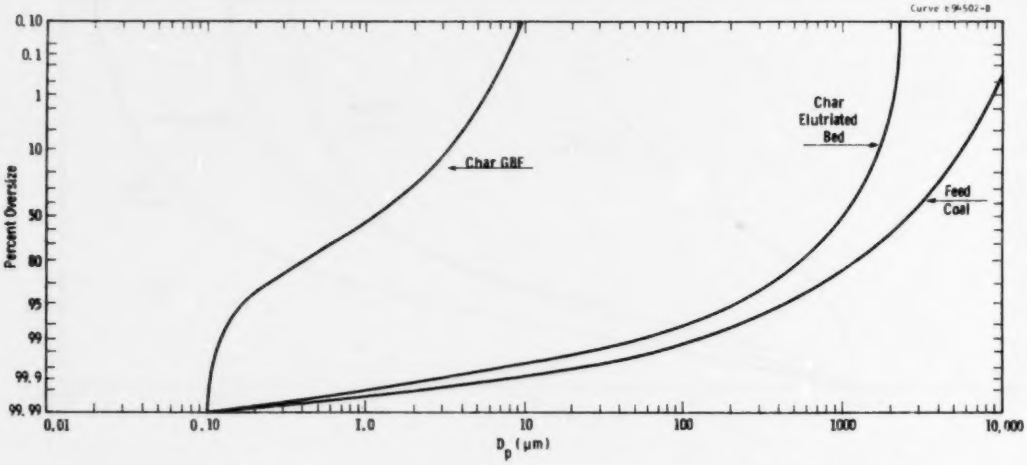

Figure D1 - Coal and char size distributions (PFBC)

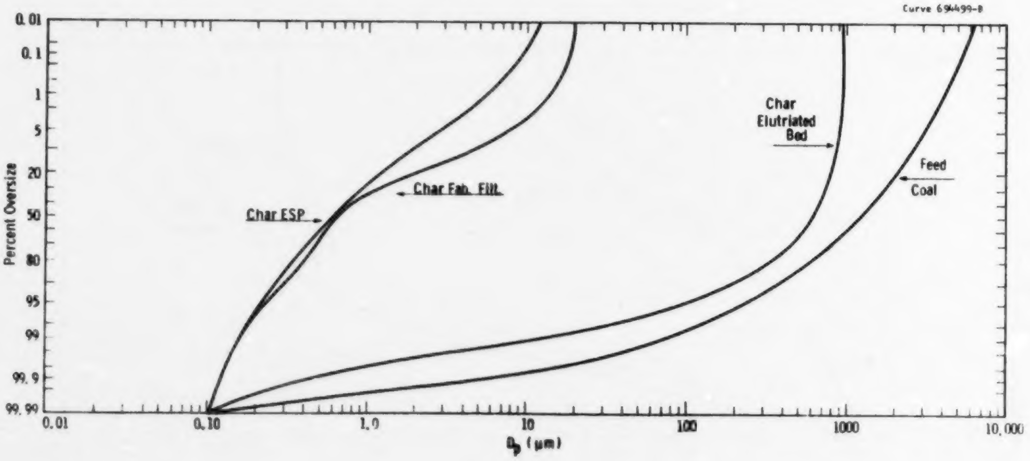

Figure D2 - Coal and char size distributions (AFBC)

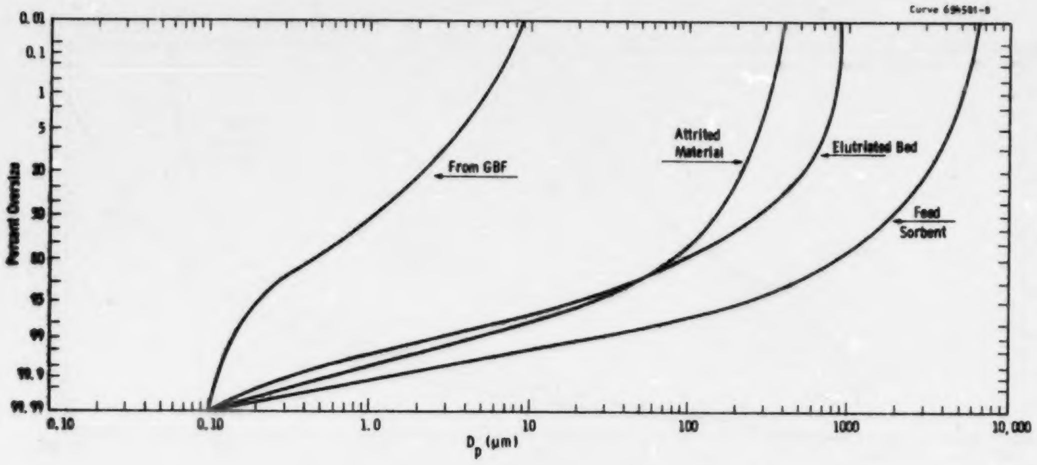

Figure D3 - Sorbent size distributions (PFBC)

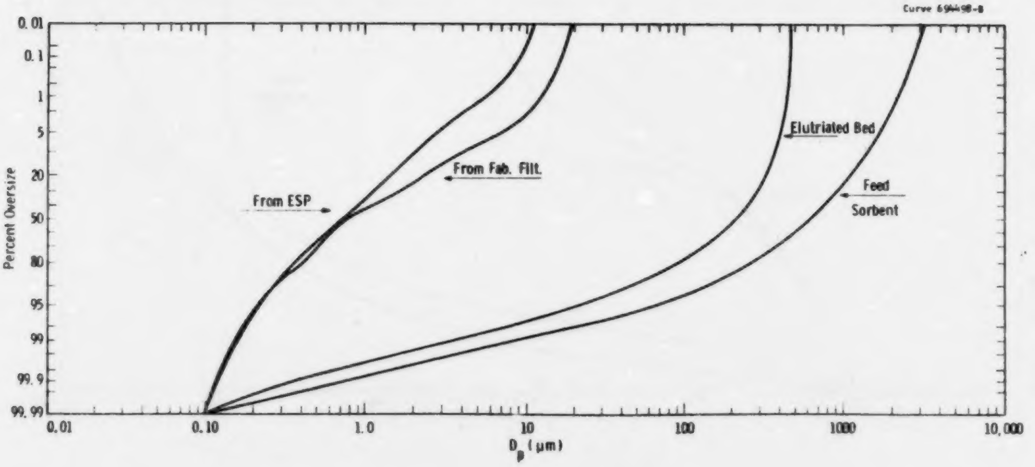

Figure D4 - Sorbent size distribution (AFBC)

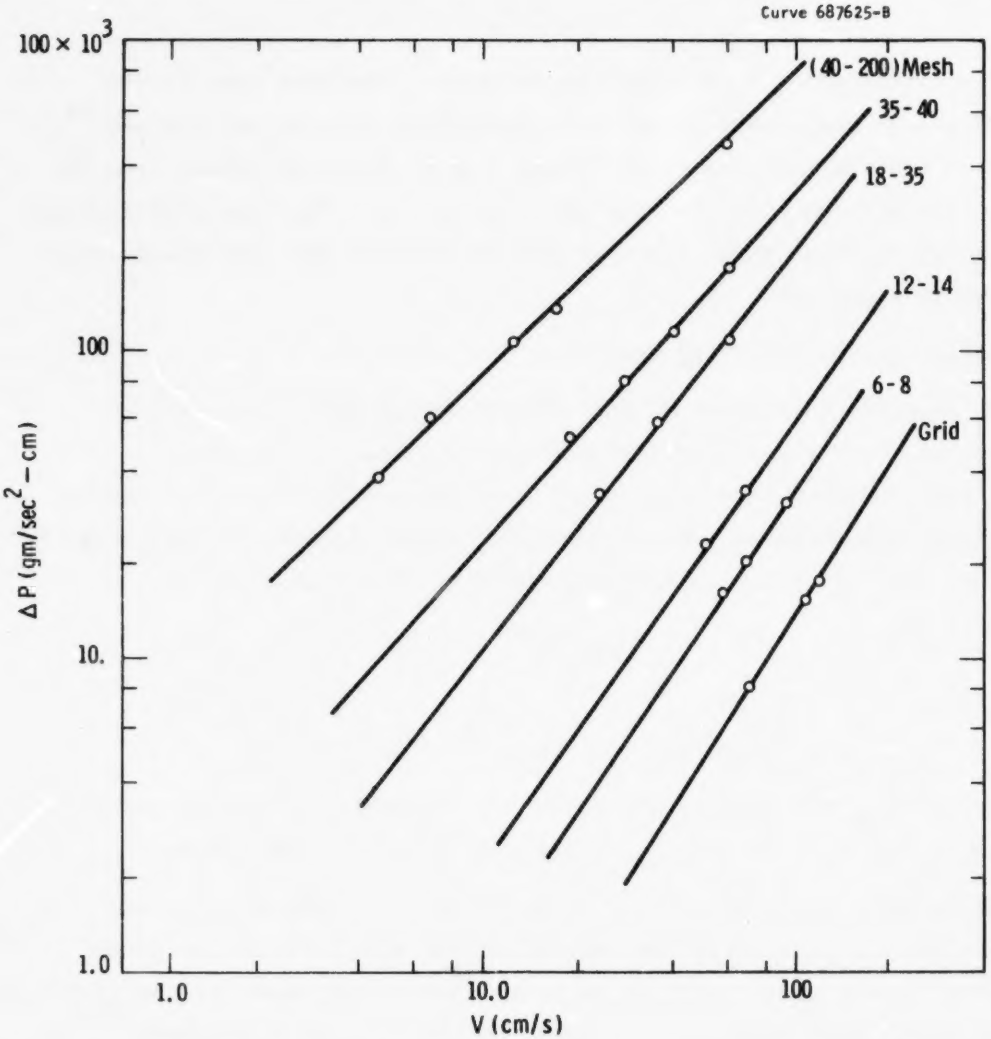

Figure D5 - Pressure drop vs. flow for different size fractions of limestone

Elutriation of Ash from the Bed

It has been assumed in this work that all the ash in the coal is elutriated, which implies that the ash elutriation rate per weight of coal fed (W_a/W_c) is equal to x_a, and the percent ash of the coal:

$$W_a/W_c = x_a \quad .$$

The size distribution of the ash was estimated from work done by the U. K. National Coal Board (NCB), and reported by Merrick and Highley.[D1] A Coulter Counter analysis of the fines from an Exxon miniplant (run 34) indicated a minimum particle size of about 0.1 μm. The size distribution constructed by these means for both the pressurized and atmospheric cases is shown in Figure D6.

Elutriation of Char from the Bed

To estimate the amount of char lost from the bed it was assumed that all of the combustion efficiency loss was due to lost char. Exxon[D2] has produced a correlation of combustion efficiency with excess air and bed temperature. This allows the weight of char lost (W_{ch}) per weight of coal fed to be obtained by the following heat balance:

$$(1 - \eta_{comb})W_c H_c = W_{ch} H_{ch}$$

or

$$W_{ch}/W_c = (1 - \eta_{comb})(H_c/H_{ch}) \quad ,$$

where H_c and H_{ch} are the heating values of the coal and char respectively, and combustion efficiency η_{comb} is obtained from the correlation.

To estimate the size distribution of the elutriated char it was assumed that the char particles maintained the same size as the parent coal particle but experienced an apparent density decrease to 0.88 g/cm^3 due to loss of volatiles. The diameter of the char particle whose terminal velocity corresponds to the superficial velocity in the bed was calculated using the data of Pettyjohn and Christianson.[D3] It was then

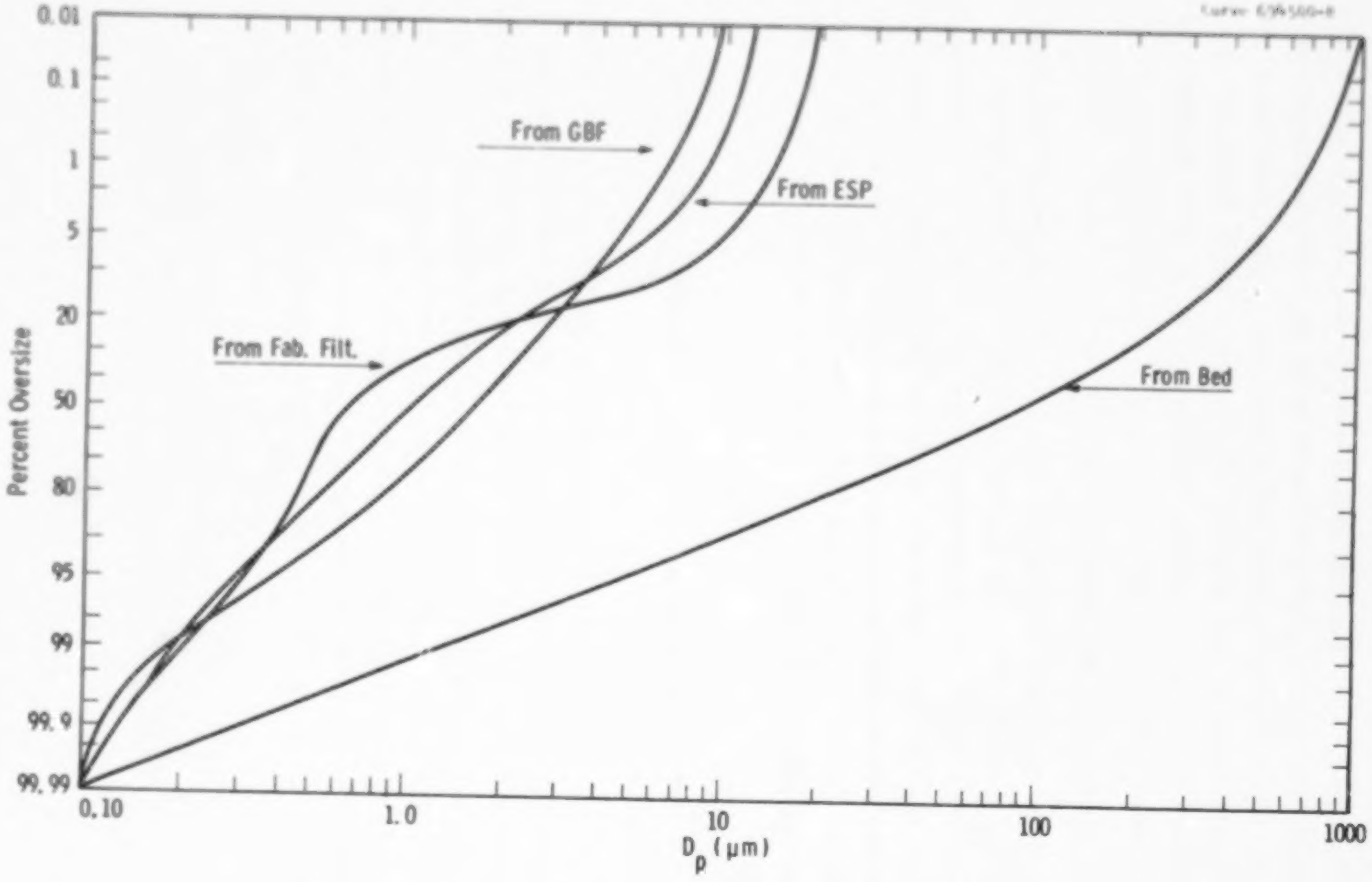

Figure D6 - Size of ash elutriated from bed for both PFBC and AFBC

assumed that the size distribution of the coal feed below the maximum diameter elutriated typified the elutriated char size distribution. Clearly, the fine fractions of the material will be elutriated faster, which will bias the distribution to the finer sizes; these finer char particles, however, are more apt to be burned, which would bias the distribution toward the coarser material. The estimate adopted here lies between these extremes.

Elutriation of Sorbent from the Bed

Sorbent elutriated from the bed can come from two sources, elutriation of fines (W_{se}) in the feed material (W_{sf}) and attrition of the material in the bed with subsequent elutriation (W_{sa}). We assumed that all of the feed material with a diameter smaller than that whose terminal velocity corresponds to the superficial velocity in the bed was elutriated immediately. If y_m is the cumulative fraction of the sorbent feed less than this critical diameter, the loading from elutriated feed is simply:

$$W_{se}/W_{sf} = y_m \quad .$$

The size distribution of this material is obtained from the feed distribution.

It is somewhat more difficult to estimate loadings and size distributions for the attrited material, since the experimental data are limited and have not been assimilated to develop predictive capability. Preliminary estimates from Exxon's miniplant and Argonne's pressurized unit indicate that roughly 4 percent/hr of the bed weight is a reasonable estimate of attrition rates. Some experimental work is currently being conducted at Westinghouse Research and Development Center[D4] that suggests that the attrition rate can be adequately described by an expression of the form

$$\text{rate} = K(U_s - U_{mf})t^{-n}$$

for the range of superficial velocities being considered here. Preliminary results indicate that the constants n and K have values of about $n = 0.7$ and $K = 3 \times 10^{-4}$ for a good attrition resistant sorbent. An average attrition rate can be calculated by integrating over the residence time (τ) for the particles.

$$\overline{X}(\tau) = \frac{1}{\tau}\int_0^{\tau} K(U_s - U_{mf})t^{-0.7}dt$$

or

$$\overline{X}(\tau) = 0.001(U_s - U_{mf})\tau^{-0.7} \quad .$$

The loading of attrited material (W_{sa}) can then be expressed by the following

$$(W_{sa}/W_{sf}) = (1 - y_m)\tau\overline{x}(\tau) \quad .$$

Combining the attrited material with the immediately elutriated material gives the total sorbent loading to the flue gas (W_{st}) as

$$\frac{W_{st}}{W_{sf}} = \frac{W_{se} + W_{sa}}{W_{sf}} = y_m + (1 - y_m)\tau\overline{x}(\tau) \quad .$$

The residence time for the solids can be estimated by the use of the following expression:

$$\tau = \frac{(288)\ (T)\ (D)\ (1 - V_f)\ (\rho'_{sorb})\ [(W_a/W_c)_{st}\phi_A + (1 - x_{ash})]}{(EF)\ (P)\ (MW)\ (U_s)\ (W_{sf}/W_{su})_{st}\ (R)\ (W_{su}/W_c)}$$

where

T = bed temp., °C
D = bed depth, cm
V_f = void fraction, including tubes
ρ'_{sorb} = bulk density sorbent, 1.38 g/cm^3
W_a/W_c = air to coal ratio - stoichiometric, 10.413
ϕ_a = air equivalence ratio, 1.20
x_{ash} = weight fraction of ash in coal, 0.107
EF = expansion factor, 1.4

P = pressure kPa, 101.3, 1013
MW = molecular weight of flue gas
U_s = superficial velocity, cm/s
$(W_{sf}/W_{su})_{st}$ = weight of sorbent to weight of sulfur, 5.75
R = calcium to sulfur ratio, 2, 5, 10, 20
(W_{su}/W_c) = weight fraction sulfur in coal, 0.043.

The pressurized case has a residence time that varied from about 2.8 to 0.28 hr when calculated in this manner, depending upon the sorbent feed rate, while for the atmospheric case τ varied from 6.8 to 0.68 for the different values of R.

The attrited material size distribution was estimated using Westinghouse experimental data[D4] as a guide. The size distribution of the combined attrited and elutriated sorbent is shown in Figures D3 and D4. We observed that for the size distributions assumed, the distribution fines created by attrition closely resembled the size distribution of the feed elutriated from the bed. This allowed one size distribution to be used for all values of R, since the relative amounts of material attrited and fines in the feed had little impact on the combined size distribution.

To obtain the mass of flue gas per gram of coal fed (W_{fg}/W_c), a mass balance around the fluidized bed was employed, assuming the following flows:

coal + air + sorbent = flue gas + char loss + ash
+ spent sorbent + half-calcined sorbent.

If R is taken as the calcium to sulfur atomic ratio used, and ξ is the percent sulfur removal (assumed to be 0.90), the overall stoichiometry can be written as

$$R(M_gCO_3 \cdot CaCO_3) + (SO_2) + 1/2\ \xi(O_2) \rightarrow (R - \xi)\ MgO \cdot CaCO_3 + \xi(MgO \cdot CaSO_4) + (R + \xi)\ CO_2 + (1 + \xi)\ (SO_2)\ .$$

This allows the mass balance to be completed. Letting

M_S = molecular weight sulfur = 32

M_D = molecular weight dolomite = 184

$M_{D/2}$ = molecular weight half calcined = 140

M_{SD} = molecular weight sulfided = 186

XS = excess air

the mass of flue gas per unit weight of coal fed is:

$$(W_{fg}/W_c) = 1 + (W_a/W_c)_{st}\ (1 + XS) - (1 - \eta_{comb})\ (H_c/H_{ch})$$
$$- X_{ash} + \frac{x_{su}}{M_s}\ [RM_D - (R - \xi)M_{D/2} - \xi M_{SD}]\ .$$

It is now possible to estimate the total loading of solids from the bed as well as the relative amounts of ash, char, and sorbent:

$$\frac{W_{total}}{W_{fg}} = \frac{x_{ash} + (1 - \eta_{comb})\ (H_c/H_{ch}) + R(W_{sf}/W_{su})_{st} x_{su} (y_m + (1 - y_m)\ \overline{x})}{(W_{fg}/W_c)} .$$

Table D1 summarizes the results of the application of these equations for both the pressurized and atmospheric cases.

Although it is difficult to make comparisons between different systems, the projected overhead loading of 0.395 g solids/g coal fed for a calcium to sulfur ratio of 2 seems reasonable, as Exxon[D5] has reported a range from 0.13 to 0.21, and ANL[D6] has reported numbers from about 0.1 to 0.5 for the pressurized case.

For the atmospheric case, total solids loadings of between 0.5 and 0.85 g solids/g coal fed have been reported[D7] at superficial velocities near the 3.05 m/s (10 ft/s) assumed here, which compares well with the value of 0.535 projected. Data on the size distributions of the solids carry-over from the combustor are scarce, but even less data on the distribution of the fines exist. Figure D7 shows the combined solids size distribution projected by this analysis for calcium/sulfur = 2 for

Curve 697242-A

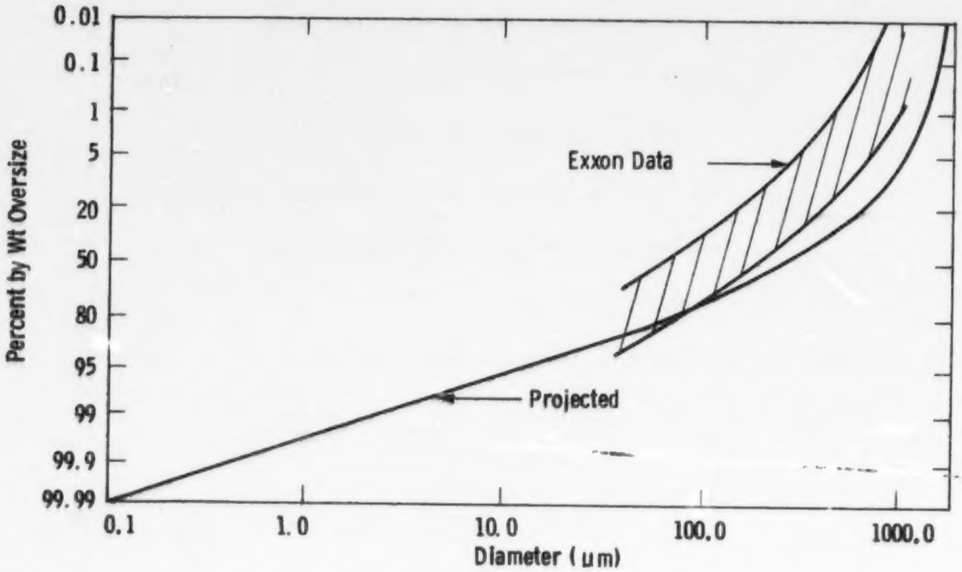

Figure D7 - Total solids from combustor (PFBC)

Curve 697243-A

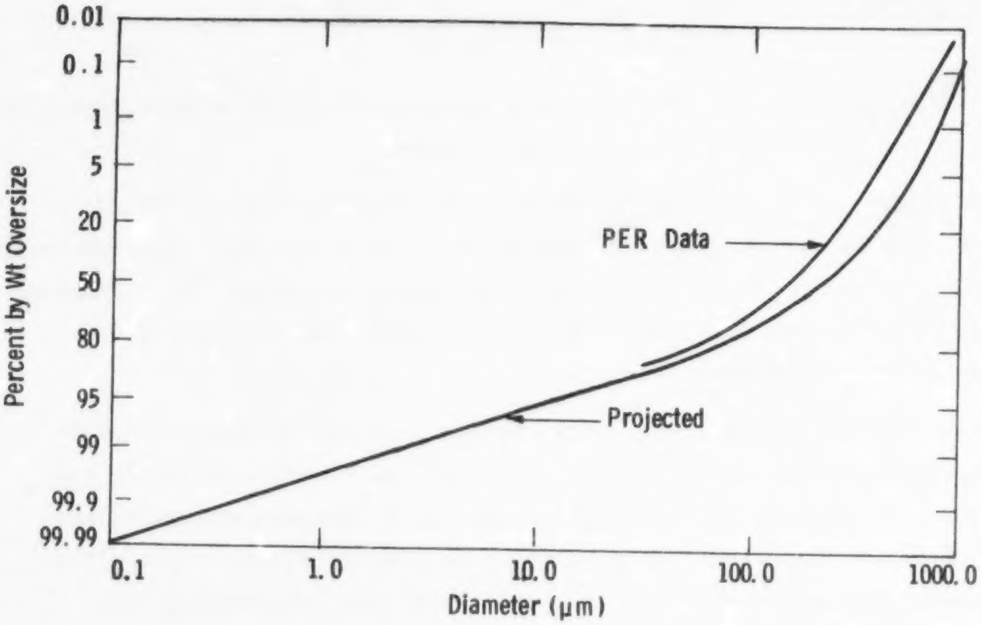

Figure D8 - Total solids from combustor (AFBC)

PFBC. Included is a range of distributions reported by Exxon.[D5] The manner in which the data are extrapolated into the fines region has a great impact on the performance of the particulate removal train. As mentioned before, particle size analysis done on Exxon fly ash indicated that the minimum particle size was about 0.1 μm. The total solids size distribution from the AFBC is shown in Figure D8. As can be observed, the projected distribution agrees rather well with the experimental data reported by Pope, Evans and Robbins.[D8]

Table D1

EFFLUENT LOADINGS FROM BED

Pressurized Case

R(Calcium/ Sulfur	g Sorb/ g Coal	g Ash/ g Coal	g Char/ g Coal	g Solids/ g Coal	ng/J
2	0.216	0.107	0.0716	0.395	14,130
5	0.472	0.107	0.0716	0.651	23,290
10	0.863	0.107	0.0716	1.042	37,280
20	1.597	0.107	0.0716	1.776	63,540
Atmospheric Case					
2	0.356	0.107	0.0716	0.535	19,140
5	0.792	0.107	0.0716	0.971	34,740
10	1.466	0.107	0.0716	1.645	58,850
20	2.749	0.107	0.0716	2.928	104,760

PROJECTIONS OF PARTICLE CONTROL DEVICE PERFORMANCE

The particulate control system envisioned for use in both the pressurized and atmospheric case is shown schematically in Figure D9. Both cases are essentially the same except for the final stage of particulate removal. A granular bed filter has been used as a basis for performance and costing for the pressurized system. Two devices have been considered for use for the atmospheric case. The first device considered was an electrostatic precipitator, the second a fabric filtration system.

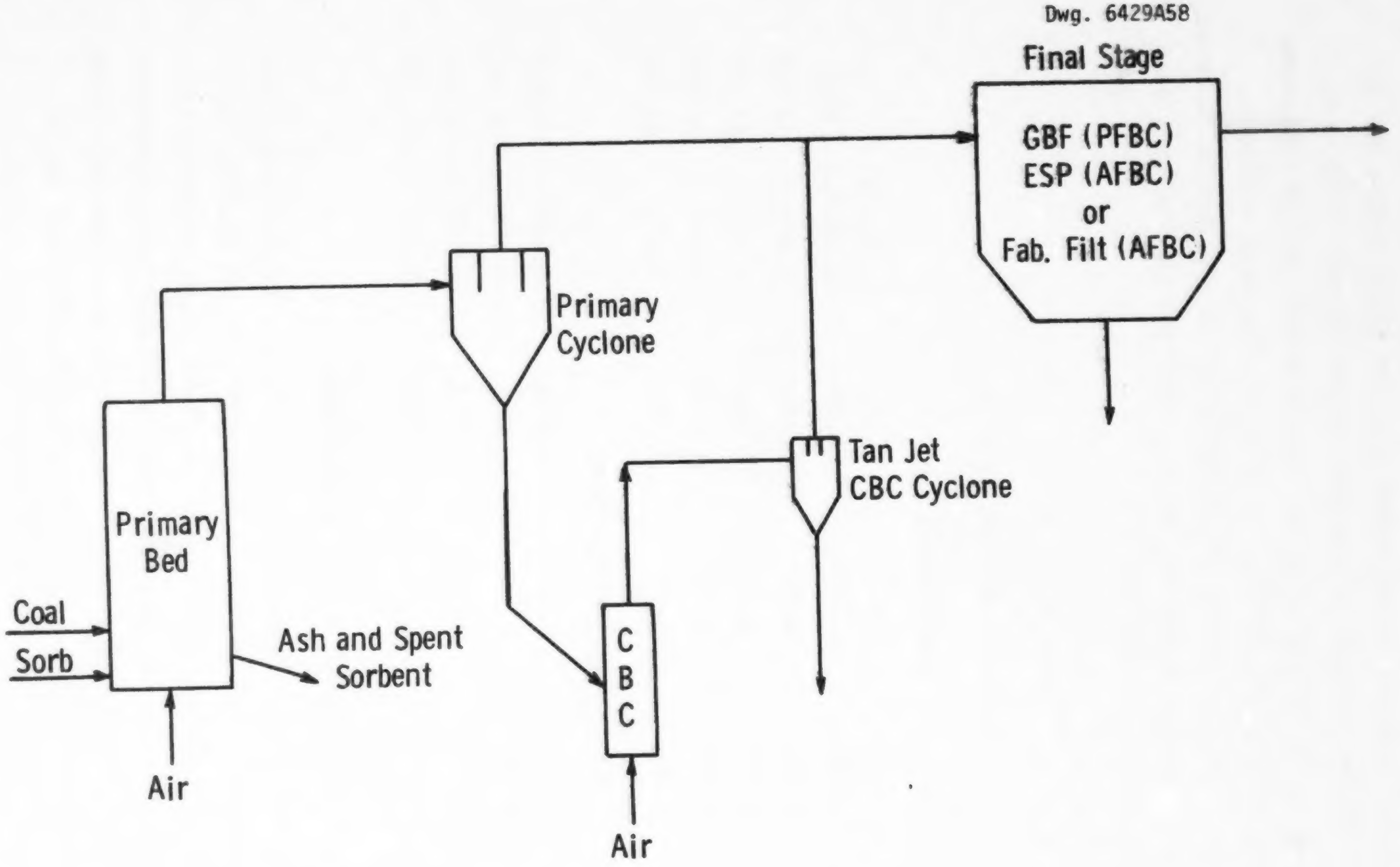
Dwg. 6429A58
Final Stage
GBF (PFBC)
ESP (AFBC)
or
Fab. Filt (AFBC)
Primary Cyclone
Tan Jet
CBC Cyclone
Primary Bed
Coal
Sorb
Ash and Spent Sorbent
Air
C
B
C
Air

Figure D9 - Schematic of particle control system

Primary Cyclone

The primary cyclones used for this study were of a fairly conventional design operating at an inlet velocity of about 24 m/s (80 ft/s). Each unit has a capacity of about 27,184 m^3/hr and a pressure drop of about 21 kPa (3 psi). A plot of particle penetration versus particle size for the various particulate densities is shown in Figure D10. These data are based on manufacturers' projected performance at pressure and temperature.

The projected performance of this cyclone on the individual species of particulates carried out of the bed is summarized in terms of overall efficiencies in Table D2.

Table D2

PRIMARY CYCLONE EFFICIENCIES

	PFBC	AFBC
Ash	0.925	0.926
Char	0.994	0.998
Sorbent	0.989	0.991

The estimated overall efficiency for all particulate matter has been calculated for the various calcium to sulfur ratios considered and is presented in Table D3. Figures D11 and D12 show the projected size distribution of the combined solids leaving the primary cyclone for a calcium to sulfur ratio of 2.

Table D3

OVERALL EFFICIENCIES OF PRIMARY CYCLONE

Calcium/Sulfur	PFBC	AFBC
2	0.972	0.977
5	0.979	0.982
10	0.983	0.985
20	0.984	0.987

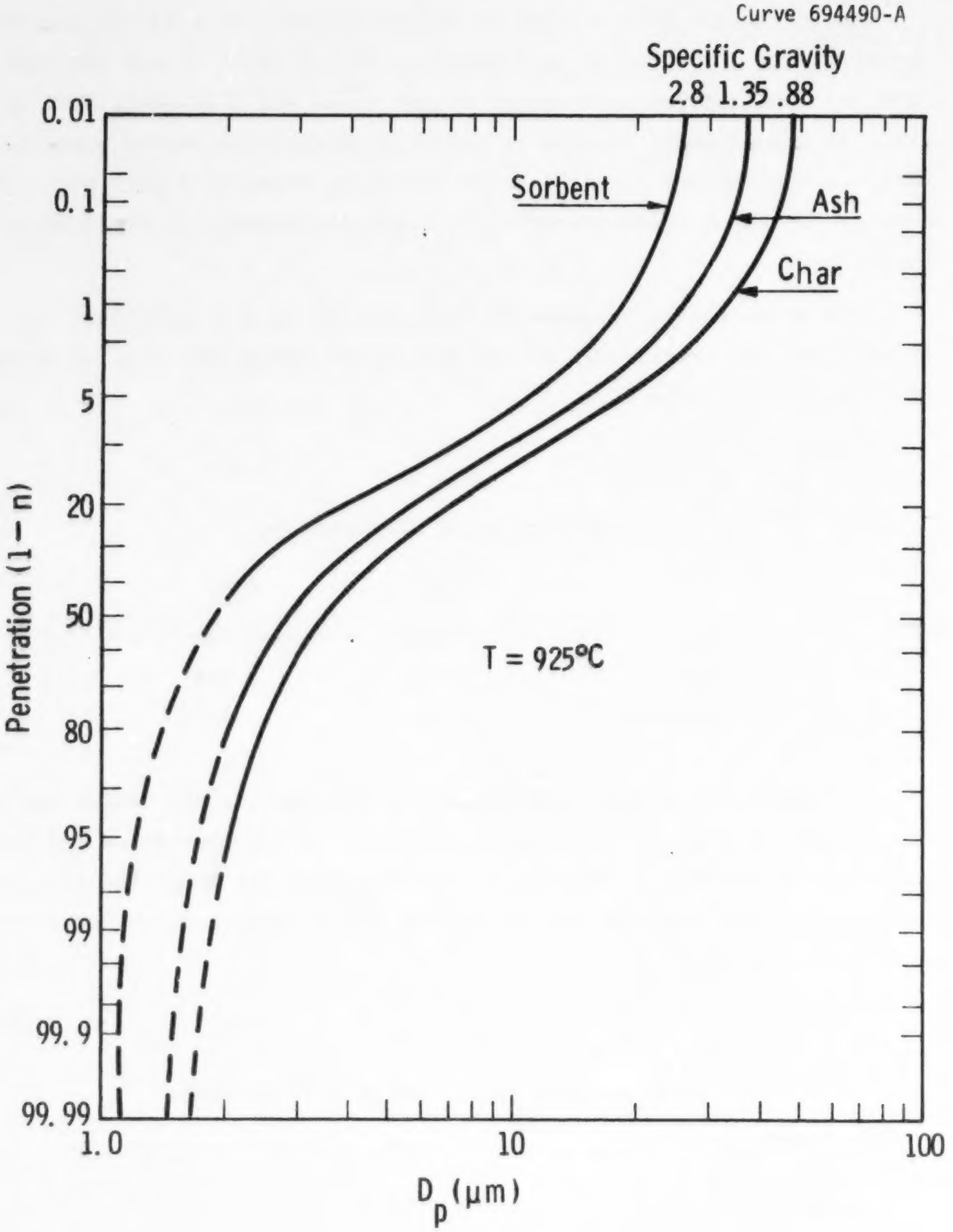

Figure D-10 - Grade efficiency for primary cyclone

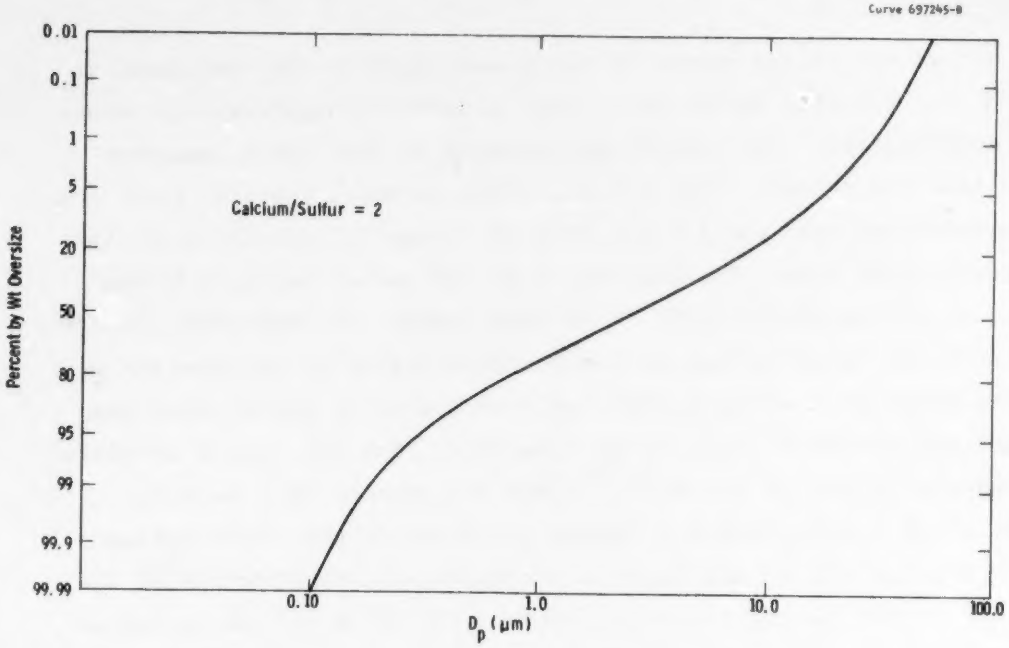

Figure D11 - Combined solids leaving primary cyclone (PFBC)

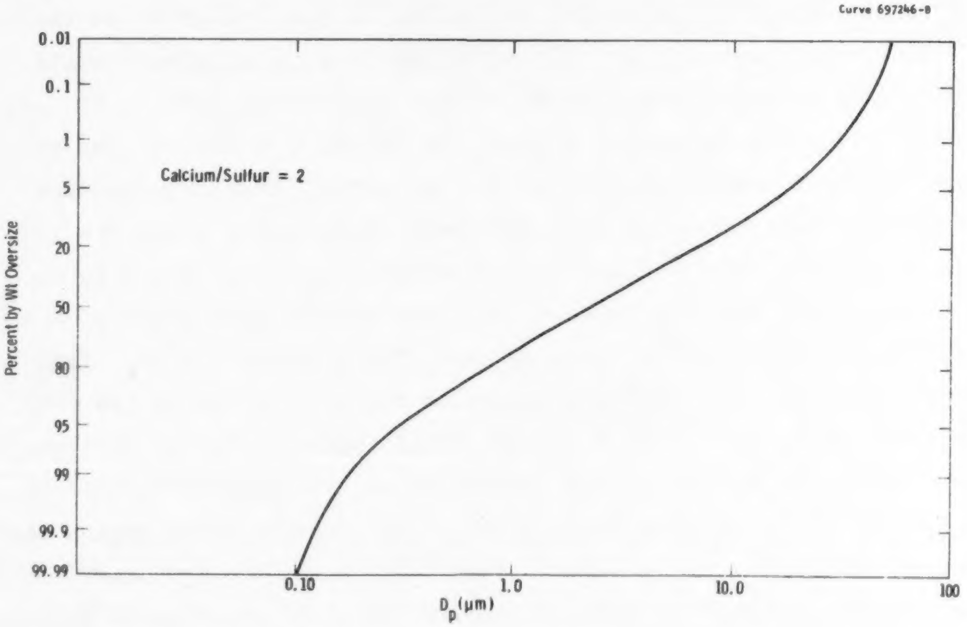

Figure D12 - Combined solids leaving primary cyclone (AFBC)

Carbon Burnup Cell

All of the solids caught by the primary cyclone are considered to be fed to the carbon burnup cell (CBC) to maintain high overall combustion efficiencies. The superficial velocity in this bed is somewhat lower than the primary - 1.7 m/s (5.5 ft/s) versus 2.1 m/s (7 ft/s) for the pressurized case and 2.7 m/s (9 ft/s) versus 3.1 m/s (10 ft/s) for the atmospheric case. The temperature of the carbon burnup cell has been taken to be 1093°C (2000°F) for both cases. The materials carried over from the carbon burnup cell were characterized in the same way as was the material from the primary bed except that no attrition of the sorbent was accounted for. It has been noted that the rate of attrition decreases with time as the sharp corners are rounded off, so it is probably not a bad estimate to ignore attrition in the carbon burnup cell. Figures D13 and D14 show the projected size distribution of the combined solids leaving the carbon burnup cell for a calcium to sulfur ratio of 2.

Carbon Burnup Cell Cyclone

Since the amount of carbon fed to the CBC is small relative to the other solids, the amount of air fed to the CBC is also relatively small. The grain loading coming from the CBC is thus rather high (200 to 400 g/m^3) [100-200 g/scf], so a nonfouling type of cyclone such as the Tan Jet by Donaldson has been considered for use at this point. Donaldson has provided estimated performance at elevated temperature curves which are shown in Figure D15. The overall Tan Jet efficiency is projected to be very high since all the particulates that pass into it have already been captured once by a standard cyclone and are, thus, fairly coarse. The cleaned gas from the CBC is then merged with the gas cleaned by the primary cyclone and is fed to the third and final stage of fine particulate removal. Tables D4 and D5 present the estimated efficiencies for the Tan Jet on the individual species and overall solids removal efficiencies as a function of calcium to sulfur ratio. Figure D16 shows the projected combined solids size distribution leaving the Tan Jet, and Figures D17

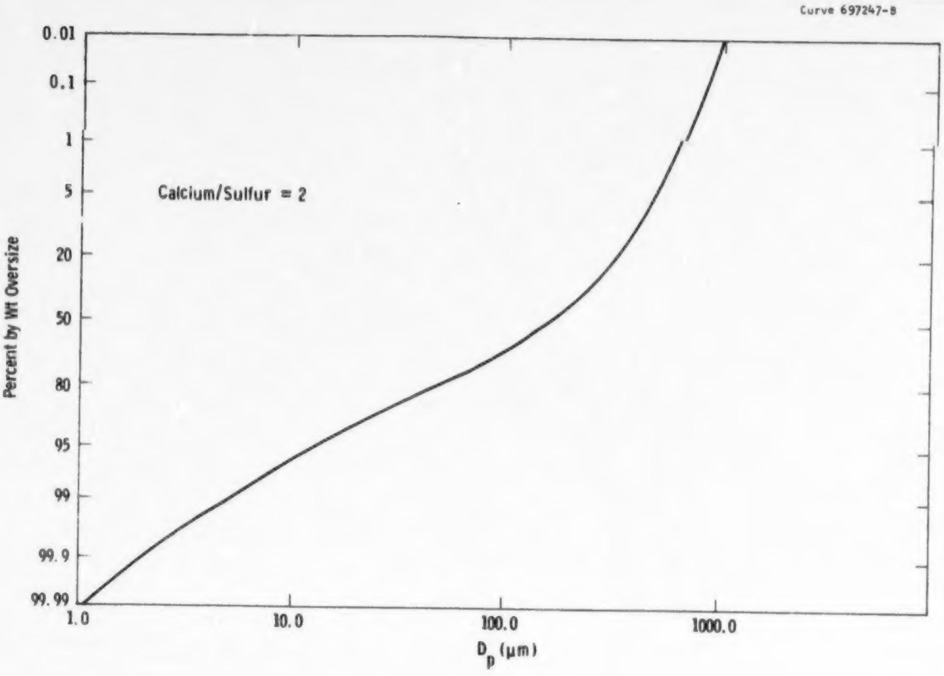

Figure D13 - Combined solids leaving CBC (PFBC)

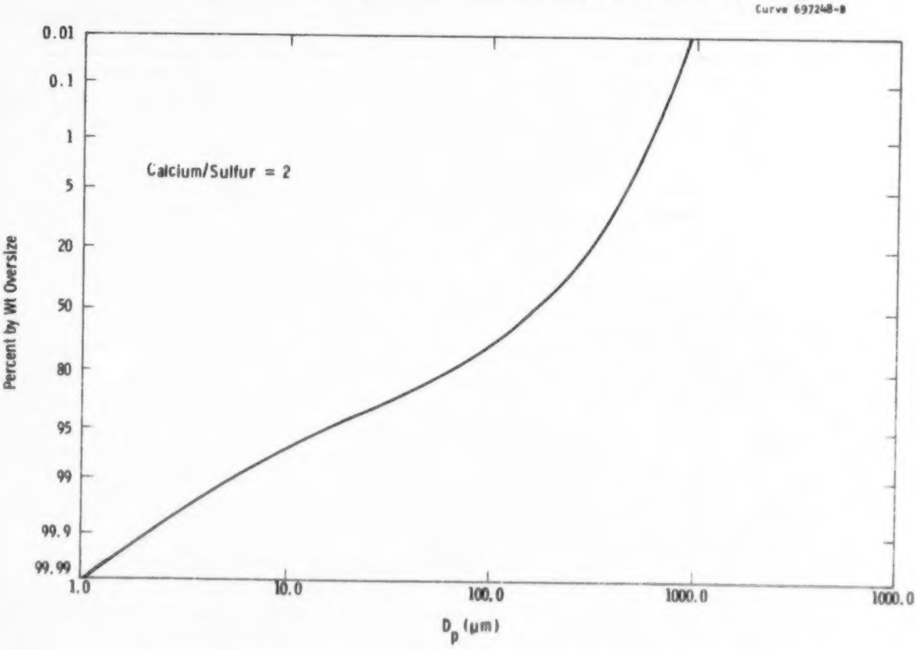

Figure D14 - Combined solids leaving CBC (AFBC)

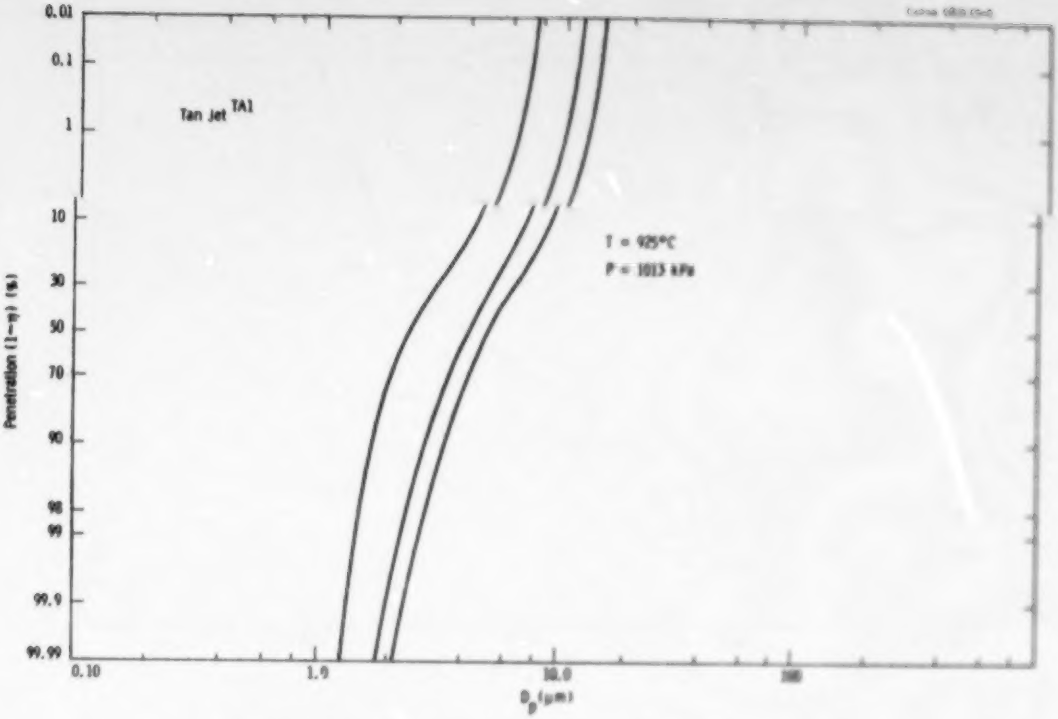

Figure D15 - Grade efficiency for fan jet

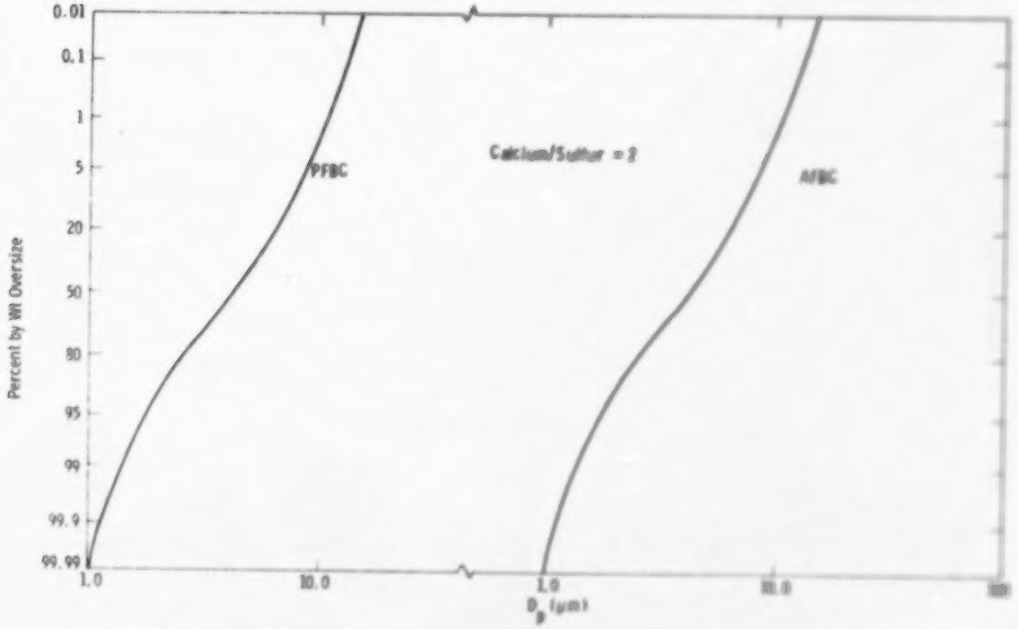

Figure D16 - Combined solids leaving the CBC cyclone

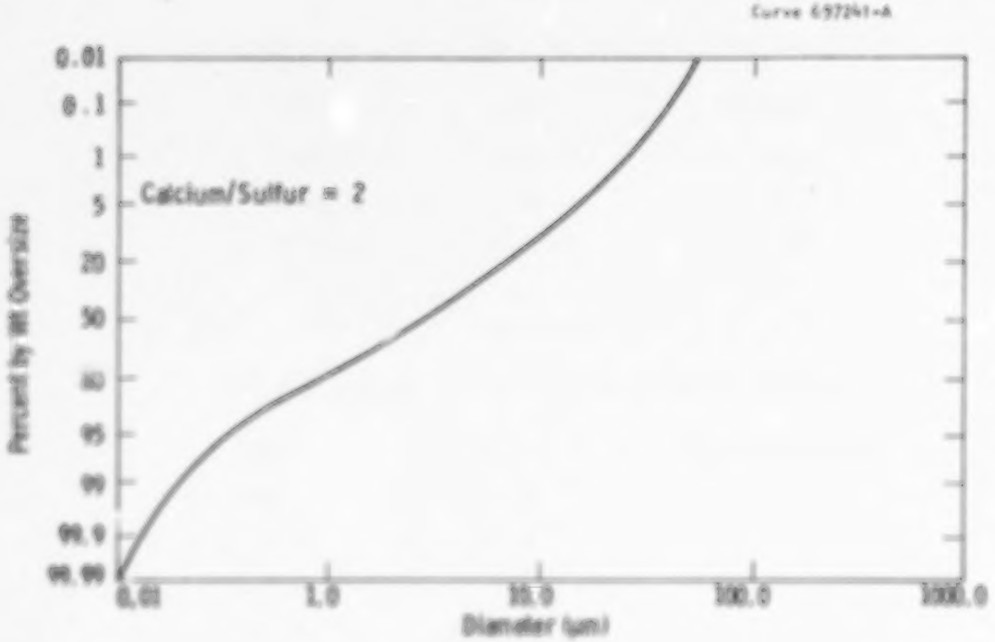

Figure D17 - Combined solids to final control device (PFBC)

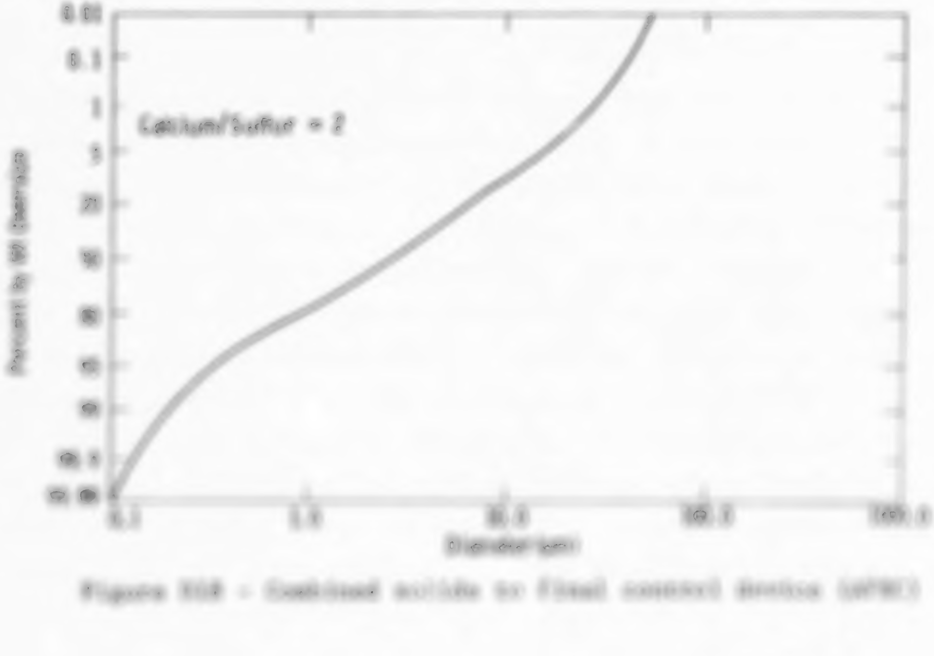

Figure D18 - Combined solids to final control device (AFBC)

Table D4

PROJECTED EFFICIENCIES OF TAN JET FOR EACH SPECIES

Species	PFBC	AFBC
Sorbent	0.9968	0.9975
Char	0.9995	0.9990
Ash	0.9769	0.9769

Table D5

PROJECTED OVERALL EFFICIENCIES FOR THE TAN JET

Calcium/Sulfur	PFBC	AFBC
2	0.9896	0.9930
5	0.9929	0.9952
10	0.9943	0.9961
20	0.9934	0.9967

and D18 show the size distribution of the merged streams from the primary collector and the Tan Jet. This stream is then fed to the final stage of particle removal equipment.

Fine Particulate Removal

It is in the choice of the fine particulate removal that the significant differences between the gas cleaning systems arise. Two major choices for the final stage of particulate removal appear to exist for the atmospheric-pressure case. These are: fabric filtration and electrostatic precipitation, and there is some degree of uncertainty in the use of either of these devices. The precise nature of the electrical properties of the ash, sorbent, and char from the fluidized-bed combustion process are not well known, so migration velocities and possible reentrainment problems in electrostatic precipitators can only be estimated. In view of the considerable problems experienced in trying to precipitate low-sulfur Western coals in conventional boilers, these considerations may be of some consequence in the fluidized-bed case. For the purposes of this work, data reported by J. D. McCain[D9] have been used. These data were obtained by measurements on a cold-side pilot-scale ESP operating on a coal-fired boiler burning low-sulfur coal. It has been assumed that the different components of the particulate are collected with equal efficiency since the grade efficiency curve used already indicates the problems of reentrainment and gas bypassing by the lower-than-expected efficiencies in the larger particle sizes. The hot-side precipitator data from McCain[D9] do not indicate this problem with the larger particle sizes at only modestly larger collection area to flow ratios. The lower efficiency cold-side data were adopted as a conservative estimate of performance. The grade efficiency assumed for the ESP from McCain[D9] is shown in Figure D19.

The other alternative for the atmospheric case is the use of a fabric filter system. There may be some uncertainty associated with the erosiveness of the particulate, and hence with bag life, but it seems likely that operating conditions can be chosen to avoid serious maintenance problems with a baghouse system. There is probably less uncertainty about the performance of the filter system, compared to the ESP, since the physical nature of the particles is unlikely to have much impact on the capture efficiency of the filter. Baghouses, however, do have potential problems with start-up and blinding, which necessitate careful operation. The grade efficiency used in this study has been taken from data reported by H. Spagnola[D10] and is shown in Figure D20. These data were collected from a large baghouse facility operating on power plant flue gases.

The final stage of cleanup for the pressurized system is probably the least well-defined piece of equipment, since hot pressurized particulate removal is not yet state-of-the-art practice. The device that has been used as the basis of this study is a granular bed filter of the Ducon Corporation design. Since preliminary operation of a 800 m^3/hr scale granular bed filter unit at the Exxon miniplant has been plagued by various mechanical problems, data from bench-scale experiments conducted by Westinghouse[D11] have been used rather than those from the

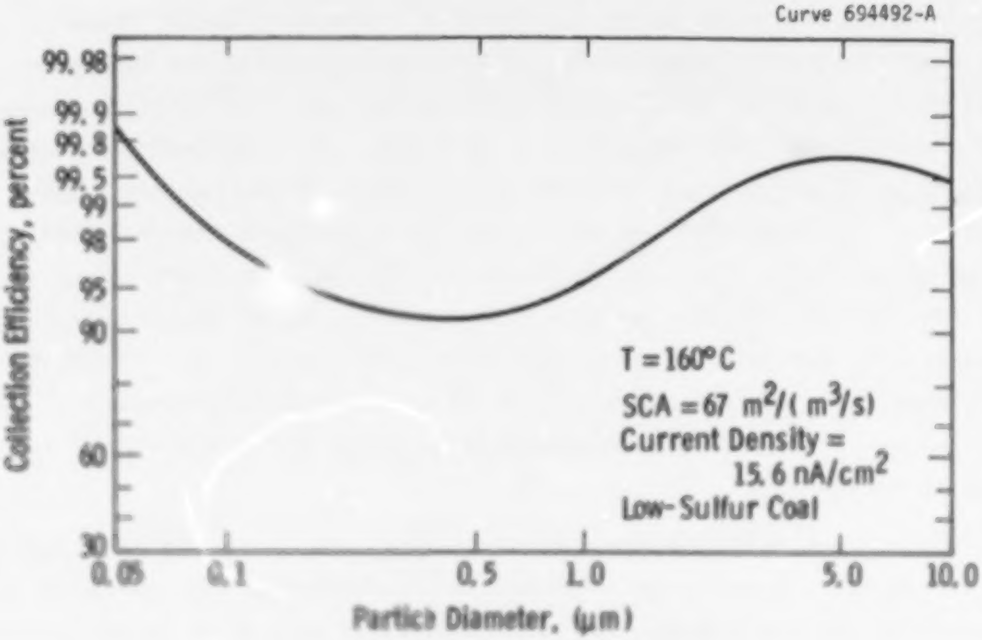

Figure D19 - Grade efficiency for ESP(Ref.D9)

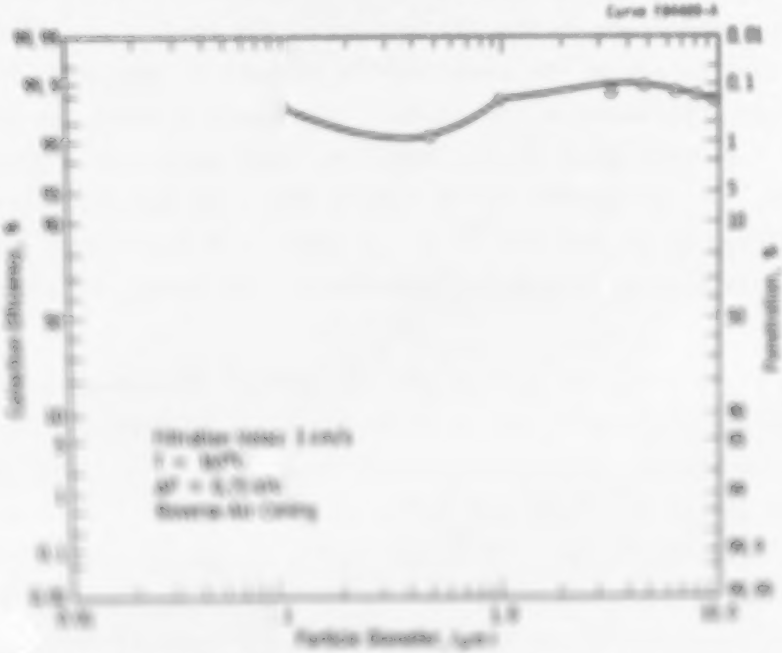

Figure D20 - Baghouse performance at Sunbury Steam Electric Station (Ref. D10)

miniplant. The grade efficiency used in this study is shown in Figure D21. It should be noted that these data were obtained on a small experimental unit and are more optimistic than those realized in larger units. Confirmation of this performance in larger units is required.

The entire hot gas fine particulate removal picture may change if the preliminary work done on rigid ceramic filters reported by Ciliberti[D12] can be developed to commercial scale. These filters may have some problems with cleaning if a sticky ash is encountered. They appear, however, to have a large cost advantage over granular bed filters as well as significantly higher collection efficiencies in the submicron range. Ceramic bag filters are also promising, as indicated by work reported by Acurex.[D13]

Results

The projected effectiveness of each control system for all three of the particulate species is shown in Table D6 where the fractional penetration for the overall system is presented.

Table D6

OVERALL FRACTIONAL PENETRATION FOR EACH SPECIES

	PFBC	AFBC	
Species	GBF	Fab. Filt.	ESP
Sorbent	0.000263	0.000030	0.000313
Char	0.000030	0.000018	0.000189
Ash	0.001116	0.000161	0.00125

These species penetrations can be weighed to give the projected overall particulate penetrations that are present in Table D7. As can be noted from the individual species size distribution curves of Figures D2 through D4 and D6, all of the particulates emitted from the final stage device for all cases have essentially the same size distribution, with the effluent being generally log-normal and running from 0.1 μm to about 10 to 15 μm, with a mass median diameter from 0.5 to 1 μm. Since

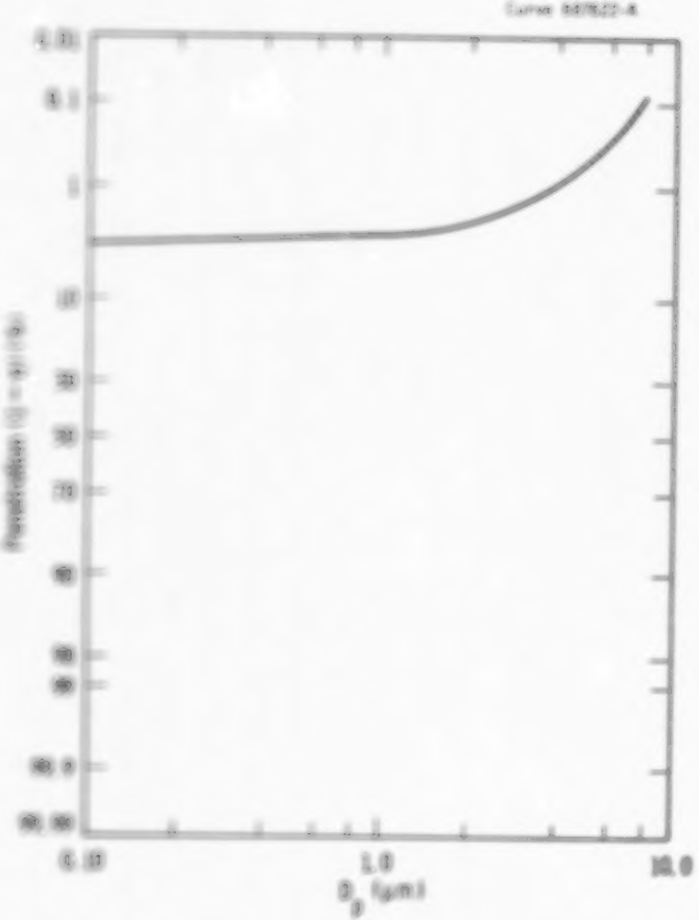

Figure D21 - Grade efficiency for granular bed filter

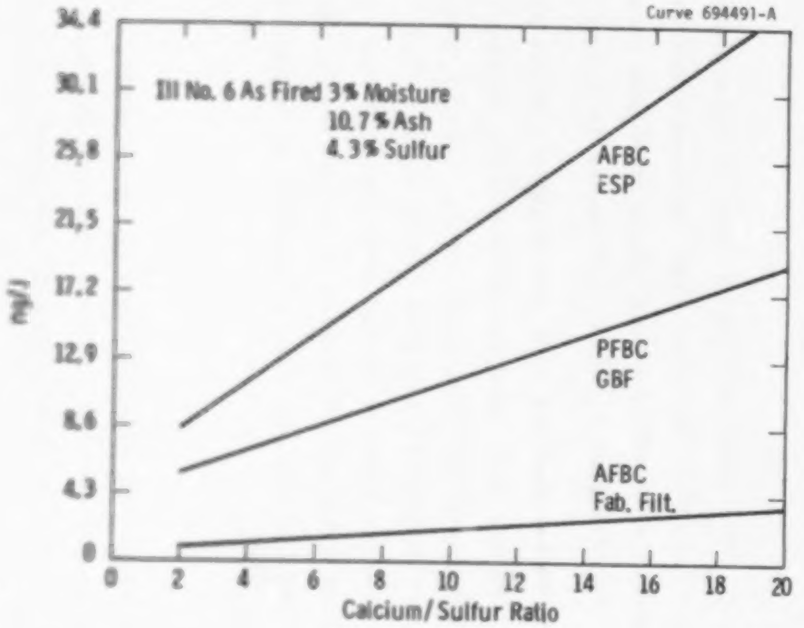

Figure D22 - Projected particulate emitted vs. calcium to sulfur ratio

the amount of sorbent fines produced by the feed and attrition vary with the calcium to sulfur ratio used, the data are presented with R as a parameter. These penetrations are projected by applying the appropriate grade efficiency curve to the mass loadings and size distribution shown in Figures D17 and D18.

Table D7

OVERALL SYSTEM PARTICULATE FRACTIONAL PENETRATION

	PFBC	AFBC	
R(Ca/S)	GBF	Fab. Filt.	ESP
2	0.000451	0.000055	0.000485
5	0.000382	0.000044	0.000408
10	0.000334	0.000038	0.000369
20	0.000305	0.000034	0.000344

Since the total loading to the system increases as R increases, the total mass escaping the system increases even though the penetration decreases slightly with increasing R as indicated in Table D7. This effect is shown in Figure D22, which is a plot of the total weight of particulate escaping the system per unit energy of coal fed.

For AFBC, the electrostatic precipitator is projected to control the emission of particulate material to below the particle emission requirement of 12.9 ng/J (0.03 lb/MBtu) for calcium to sulfur ratios less than five. The fabric filter system is projected to control the particulate emissions to levels less than 12.9 ng/J (0.03 lb/MBtu) for all calcium to sulfur ratios considered. With the investment and operating and maintenance costs of electrostatic precipitators and baghouses not being sufficiently different to have a significant impact on the overall cost of electricity, the baghouse system is recommended.

Particulate control with PFBC may be different from AFBC in the sense that the process constraints on particulates may supersede the environmental requirement of 12.9 ng/J (0.03 lb/MBtu), depending on size distribution. The granular bed filter is projected to meet the emission control target for calcium to sulfur ratios less than 12.

The projections of sorbent requirements in this report for 90 percent SO_2 removal ranged from a calcium to sulfur ratio of 2.9 to 7.0 for AFBC, and from 1.7 to 4.5 for PFBC. It is, thus, apparent that, according to these projections of particle control performance, the targets of 90 percent SO_2 removal and 12.9 ng (0.03 lb) particulate/J (MBtu) should be achievable simultaneously. As emphasized earlier, the estimates of particle control device performance used here are largely projections that must be confirmed on operating fluidized-bed combustion units.

The impact of coal sulfur-content, coal ash-content, and coal heating value are important in determining the particulate emission, but their influence should not change the conclusions already reached. The intercept of the curves in Figure D13 at a calcium to sulfur ratio of 0 should increase approximately linearly with an increase in the ratio of the coal ash-content over the coal heating value. The slope of the curves will increase approximately linearly with the quantity of coal sulfur-content over the coal heating value. As an example, a 50 percent increase in the coal sulfur-content and coal ash-content from the base values in Figure D13 (4.3 percent and 10.7 percent, respectively) would result in a particulate emission for the fabric filter system of less than 12.9 ng/J (0.03 lb/MBtu) for the range of calcium-to-sulfur ratios considered and would result in a particulate emission less than 12.9 ng/J (0.03 lb/MBtu) for granular bed filters at calcium to sulfur ratios less than 3 and for electrostatic precipitators at calcium to sulfur ratios less than about 1.5.

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APPENDIX E

CONVENTIONAL POWER PLANT DESIGN AND COST BASIS

The cost of electricity for a 796 MW conventional pulverized-coal power plant with lime-slurry wet-scrubbing for sulfur oxide control has been projected at two levels of control: 1.2 lb SO_2/MBtu (516 ng/J) (83% sulfur removal efficiency for a 4 wt % sulfur coal) and 90 percent sulfur removal efficiency. The basis for these projections is a design study performed by General Electric/Bechtel and described and expanded upon by TVA.[E1]

The GE and TVA study used a basis consistent with that applied for conceptual AFBC and PFBC power plant design (the basis applied in the ECAS program) and is thus valuable for comparison purposes.

The General Electric conventional plant design is for a 3.9 wt % sulfur coal, meeting a standard of 90 percent sulfur removal for a 4.5 wt % sulfur coal. The conventional plant has a stack gas temperature of 175°F (79°C) (50°F [28°C] of reheat) and operates with 33.8 percent plant thermal efficiency. The lime-slurry scrubber system was on-site calcination of limestone and has a 5-year slurry disposal capacity and a 5.5-year plant construction time. The wet lime absorber parameters are summarized in Table E1. The General Electric/Bechtel costs are used directly for the case of a 4 wt % sulfur coal with 90 percent sulfur removal efficiency except for two changes suggested by TVA: pond size was increased to a 30-year capaity; and the stack cost was increased to reduce stack corrosion.

The unmodified General Electric/Bechtel conventional plant cost is \$771.3/kW. TVA recommends an increased stack cost of \$4.33 x 10^6 and increased disposal cost of \$22.2 x 10^6, resulting in a total plant investment \$805.6/kW. The cost of electricity is

Capital charges	25.46 mills/kWh
O&M	3.74
Fuel	10.10
Total	39.30 mills/kWh

The cost of limestone, with a calcium-to-sulfur ratio of 1.10 and a delivered cost of $5/ton ($5.51/Mg), is included in the O&M cost at 0.374 mills/kWh.

TVA modified the General Electric/Bechtel design for the conventional power plant for the case of 83 percent SO_2 removal with a 3.9 wt % sulfur coal, using the premises listed in Table E2, with a 30-year disposal capacity and on-site limestone calcination.

The TVA-modified conventional plant electrical capacity is 798.2 MWe. The plant capital investment is $782.3/kW, and the cost of electricity is, with limestone at ($5/51/Mg).

Capital charges	24.73 mills/kWh
O&M	3.74
Fuel	10.10
Total	38.57 mills/kWh

Both of the General Electric/Bechtel and TVA conventional plants are assumed to be capable of satisfying the more stringent NO_x and particulate emission levels considered in this report with little cost impact.[E2]

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E2. Effects of Alternative New Source Performance Standards on Flue Gas Desulfurization System, Report to EPA, PEDCo Environmental, Inc., March 1978, EPA-600/7-78-033, NTIS PB 279 080.

Table E1

WET LIME ABSORBER SYSTEM PARAMETERS, CONVENTIONAL FURNACE-STEAM CYCLE

(Basis: 90% SO_x removal for 4.5% sulfur coal -
General Electric/Bechtel Design)

SO_2 Absorbers (6)	TCA-Type
No. of Stages	3 (6 in. of Spheres/Stage)
Superficial Gas Velocity	8 ft/s
Total Pressure Drop	9 in. H_2O
Liquid/Gas Ratio	72 gal/Mscf
Presaturation Sprays	2.5 gal/Mscf
Mist Eliminator Wash Sprays	2 gpm/ft^2
Lime: SO_2 Stoic. Ratio	110%
Absorber Hold Tank Residence Time	5 min
Recycle Slurry Solids	10% wt
Lime Makeup Slurry Solids	20% wt
Spent Slurry Pond Solids	40% wt

Table E2

WET LIME ABSORBER PARAMETERS, CONVENTIONAL FURNACE-STEAM CYCLE

(Basis: 83% SO_x removal for 3.9% sulfur
coal - TVA design)

SO_2 Absorbers (4)	
No. of Stages	3 (TCA-type)
Superficial Gas Velocity	12.5 ft/s
Total Pressure Drop	6 in. H_2O in the scrubber
Liquid/Gas Ratio	50 gal/Mscf
Presaturation Spray	2.5 gal/Mscf
Lime/SO_2 Absorbed Stoichiometry	1.05
Absorber Hold Tank Residence Time	10 min
Recycle Slurry Solids	8% wt
Lime Makeup Slurry Solids	20% wt
Spent Slurry Pond Solids	40% wt
Calcination Off-Gas meets SO_2 emission requirements	